

CRANFIELD UNIVERSITY

MICHELLE GANDER



**LOW-COST MEMBRANES FOR SUBMERGED MEMBRANE
BIOREACTORS**

SCHOOL OF WATER SCIENCES

MPhil THESIS

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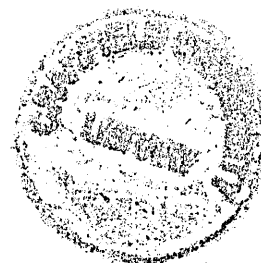
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Michelle Gander



Low-cost membranes for submerged membrane bioreactors

Supervisor Dr S. J. Judd

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Master of Philosophy**

ABSTRACT

The present work aims to assess the feasibility of using an ostensibly low-cost polypropylene (PP) material for processing municipal wastewater from the University sewage works. Two different types of hydrophilic membranes were employed, both being macroscopically homogenous (i.e. isotropic) in their physical nature. These were:

- An extruded polymeric membrane (EPP) containing microscopic glass beads as a fibrulating agent (0.5 μm pore size), and
- A non-woven felt of microscopic fibres (NWPP) of 1.5, 3 and 5 μm pore size.

The membrane materials were mounted on plate and frame modules and used in a submerged membrane bioreactor (MBR) run at 0.06 bar, fed continuously with primary effluent. Membrane performance was assessed in terms of specific flux and rejection, the latter with respect to indicator bacteria. Overall bioreactor performance was measured in terms of BOD₅ and COD removal and the extent of nitrification. An initial economic analysis was also undertaken to identify major costs in treating municipal wastewater with the two configurations of MBR. Analysis of these two areas concluded that:

- flux decline is most significant in the first few hours of filtration,
- bacterial rejection increases with the formation of the dynamic membrane,
- organics removal is high even when mixed liquor suspended solids are low,
- sludge production is reduced due to long solids retention times and substrate-inhibited microorganisms, and
- economics point to submerged MBRs being more cost efficient than side-stream MBRs due to their lower operating costs.

Only one NWPP membrane (5 μm pore size) could be concluded to be suitable for use to treat municipal wastewater, and only if disinfection was not required. The other NWPP membranes were highly susceptible to fouling, therefore producing a permeate flux unacceptable for commercial treatment processes. The PS membrane showed the

most suitable characteristics for treatment of municipal wastewater in submerged MBRs.

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CHAPTER ONE
INTRODUCTION

1.0 Background

In the UK, control of wastewater effluent discharges to watercourses is controlled by the Environment Agency, which issue, monitor and enforce the standards set by the European Union. The two most important pieces of legislation, with respect to domestic wastewater discharges, are the 1976 Bathing Water Directive and the 1991 Urban Wastewater Directive (EEC, 1976, 1991). The former stipulates mandatory standards for total coliforms (10,000/100ml), faecal coliforms, *Salmonella* and enteroviruses allowed in bathing waters. The latter stipulates levels of BOD₅ (25 mg/L), COD (125 mg/L), total suspended solids, nitrogen (10–15 mg/L) and phosphorus (1–2 mg/L) levels.

1.1 Technologies

Treatment technologies capable of removing nutrients to the current standards are Biological Aerated Filters (BAF), Trickling Filters (TF) and the Activated Sludge Process (ASP). However, the ranges with which these treatment options remove bacteria and viruses is large, e.g. ASP removal of various microorganisms is in the range of 25–99%. Therefore, if disinfection is required, a tertiary treatment process must be applied, for example gravel/sand filters, ozonation or chemical precipitation; this incurs higher capital and/or operating costs. Membrane Bioreactors (MBRs) have the advantage of not only reducing biological treatment and disinfection to one step, but of presenting a physical barrier to specific components of the wastewater, e.g. bacteria and large macromolecules requiring long retention times for bacterial breakdown. However, the high capital and operating costs of membrane systems suppressed their application to domestic wastewater treatment.

1.2 Membrane Processes for Wastewater Treatment

Membrane filtration offers a treatment option producing a good quality effluent that is largely independent of influent water quality. Other benefits include a small footprint,

the ability to process high organic loads, and reduced sludge production. The stricter standards for discharge combined with increasing costs of sewage sludge disposal make the use of membrane processes a more plausible alternative to the conventional wastewater treatment plants, notwithstanding their high capital cost.

There are a number of large-scale membrane treatment plants now being used for the treatment of municipal wastewater. The *Memcor* process comprises membranes configured as hollow fine fibres that are regularly cleaned by air backflushing. The process has been employed downstream of pre-filtration or lamellae settling at the Aberporth sewage treatment works on the Welsh coast. The *Renovexx* system comprises a woven tube onto which a precipitated coagulant membrane is formed in-situ, this coagulant layer acting as the separation barrier. The *Kubota* system utilizes flat plate membrane units that are aerated; this promotes both fluid turbulence and aerobic degradation of organic matter. The Kubota process has been installed at Porlock sewage treatment works, treating up to a 3500 population equivalent (p.e.) for effluent discharge direct to the sea.

1.3 Research Objectives

As the capital costs of membrane processes are directly related to the expense of the membrane material, which in turn is related to performance, it is of practical interest to assess the capabilities of less costly membrane materials. These inexpensive membrane materials may not be expected to give such a high effluent quality, but they may perform well enough for legislative purposes. The purpose of this research is to investigate the effects of various inexpensive membrane materials on the quality of the effluent produced, specific flux and fouling propensity.

The main objectives of this research are to ascertain:

- the specific fluxes and operational cycles attainable with simple and minimal cleaning for low-cost membrane materials,
- the disinfection properties of such membranes, by testing for indicator bacteria, and

- overall bioreactor performance by determining various parameters such as BOD₅, suspended solids and ammonia.

CHAPTER TWO

AEROBIC MBRS FOR DOMESTIC WASTEWATER TREATMENT: A REVIEW WITH COST CONSIDERATIONS

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Technology* 18(2) 119–130**

AEROBIC MBRS FOR DOMESTIC WASTEWATER TREATMENT: A REVIEW WITH COST CONSIDERATIONS

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Abstract Membrane bioreactors (MBRs) present a means of intensively biologically treating high COD or BOD wastewaters but, like other membrane processes, are constrained by their tendency to foul. Fouling is the general term given to those phenomena responsible for increasing membrane hydraulic resistance. It can be reduced by maintaining turbulent conditions, operating at sub-critical flux and/or by the selection of a suitable fouling-resistant membrane material.

The performance of various MBRs is appraised with reference to (i) fouling propensity, and (ii) removal of organics and microorganisms. Energy costs for the two process configurations for MBRs, submerged and side-stream, are reported with particular attention paid to aeration and recycle pumping costs. A number of commercial plants treating domestic wastewater are described, with further details of the most recent full-scale MBR for sewage treatment tabulated.

It is shown that the side-stream configuration has a higher total energy cost, by up to 2 orders of magnitude, compared with the submerged system due to the recycle component. The submerged configuration operates more cost effectively than the side-stream configuration with respect to both energy consumption and cleaning requirements, with aeration providing the main operating cost component as it is required for both mixing and oxygen transfer. On the other hand, the lower flux under which the submerged system operates implies a higher membrane area and thus a higher associated capital cost. It is concluded that the MBR is a highly effective treatment process for wastewater treatment in areas requiring a high quality effluent (such as

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discharge to bathing waters or water reuse) or specialisation in the microbial community (e.g. high strength liquors, effective nitrification).

Keywords Biological removal; commercial units; energy costs; fouling remediation; membrane fouling; bacteria and virus rejection

1 Introduction

Current and impending legislation in the EC on wastewater effluent discharge has led to the need for enhanced treatment processes capable of removing high percentages of BOD, suspended solids, nitrogen, phosphorus and bacteria [1,2]. One of the most promising newer technologies is the membrane bioreactor (MBR), a process involving membrane filtration combined with biological treatment. In an MBR solid/liquid membrane filtration occurs either within the bioreactor (submerged configuration, Fig. 1A) or externally through recirculation (Fig. 1B) subject to a pressure drop across the membrane generated by either hydraulic head or a fitted pump. The ultrafiltration (UF) or microfiltration (MF) membrane separates the retained (or *rejected*) materials, principally solids and including microorganisms, from other material largely through sieving, allowing the water and most solute species to pass through the membrane.

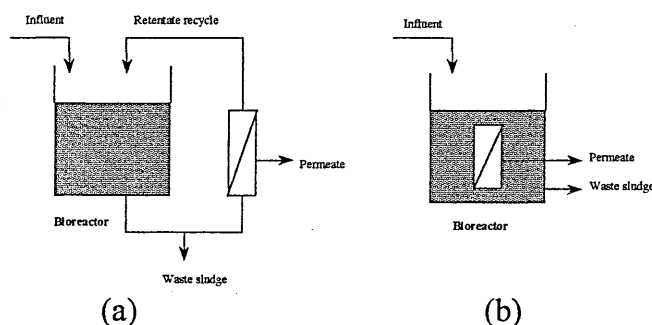


Figure 1 (a) Side-stream MBR with a separate filtration unit with retentate recycled back to bioreactor; and (b) submerged MBR: filtration unit integrated into the bioreactor

The advantages offered by MBRs over the conventional activated sludge process (ASP) include a small footprint and reduced sludge production. Submerged MBRs take only half the land area of a conventional ASP and sludge production is similarly

approximately halved (Table 1) [3]. Since sewage sludge disposal contributes significantly to overall operating costs, there are significant potential benefits in reducing its production, which would otherwise be expected to increase by 10 million tonnes dry solids per year by the year 2005 [4]. A small footprint is achievable as clarification is achieved in a single filtration stage in place of the conventional multi-stage process (primary settlement, secondary biological treatment plus possible tertiary disinfection/polishing).

Table 1 Sludge production for various wastewater treatment processes [3]

Treatment process	Sludge production (kg kgBOD^{-1})
Submerged MBR	0–0.3
Structured media biological aerated filter (BAF)	0.15–0.25
Trickling filter	0.3–0.5
Conventional activated sludge	0.6
Granular media BAF	0.63–1.06

2 Membrane Performance

The performance of a given membrane process is represented by:

- the rejection, normally expressed as a function of the ratio of the respective concentrations of the target contaminant in the feed and the permeate product (or filtrate), and
- the specific permeate flux (flux per unit pressure) or permeability, sometimes expressed as the hydraulic resistance (inverse product of specific flux and viscosity), and the transient behaviour of this parameter.

2.1 Rejection

Rejection is the removal of particles including, biological and non-biological colloids and macromolecules, by sieving or adsorption. One of the main drivers for MBR technology is the ability of the systems to disinfect, resulting in an effluent free from pathogenic microorganisms. This is an important consideration when discharge is to

bathing waters [1], or if the water is to be reused. Rejection of both bacteria and viruses by both MF and UF membranes, is known to be significant (Table 2) [7,8,9,10].

Table 2 Microorganism rejection

Membrane	Pore size/MW cut-off ($\mu\text{m}/\text{d}$)	Average log reduction	Bacteria/virus	Reference
Memtec [†]	0.2	ND	TC	[7]
PE*	0.1	4–6	Coliphage Q β	[8]
PS*	0.5	5	TC	[15]
PS*	0.3	8 [‡]	TC	[42]
Memcor [†]	0.2	3.8	FC	[52]
Renovexx [†]	0.5–1.5	3.3	FC	[52]
Stork [‡]	0.05–0.2	2.5	FC	[52]
Starcosa [‡]	0.2	8 [‡]	TC	[52]
DOW [‡]	0.2	<7	TC	[52]

* Activated sludge within MBR; [†]primary effluent; [‡]secondary effluent; [‡]indicates zero bacteria in permeate; ND, non detectable; TC, total coliforms; FC, faecal coliforms.

Chiemchaisri *et al.* [8] showed that membranes with 0.03 and 0.1 μm pore size attained the same log reduction of coliphage viruses, and that improved rejection occurred with time due to the build-up of the dynamic membrane. Madaeni *et al.* [9] showed *poliovirus* removal to be enhanced by the addition of *E.coli* to the viral culture in an MF process, possibly due to increased partial pore blocking (0.22 μm) but more likely to be caused by adsorptive retention the virus onto the larger *E. coli* bacteria.

2.2 Permeate flux

2.2.1 Permeate flux decline

Permeate flux decline is influenced by a number of factors relating to the feed water (composition), the membrane (element geometry/configuration, area and material composition), and operation (hydrodynamics). It is critically determined by the tendency of the membrane to be fouled by feed water components of the due to their accumulation on the internal and external structures of the membrane [5, 6]. This increases the overall resistance to filtration [6], thereby commensurately increasing the energy demand. The requirement for cleaning adds to the overall cost, as does membrane replacement in cases where cleaning fails to produce adequate flux recovery.

Fouling most commonly takes place external to the membrane, forming a dynamic layer at the membrane surface. As most membrane processes operate in the crossflow mode, fouling through the formation of such a dynamic layer might be expected to reach equilibrium once the adhesive forces between the layer and the membrane substrate are balanced by the shear forces at the layer:solution interface. In practice, equilibrium is not always attained, indicating some component of the overall hydraulic resistance to be time dependent.

The flux decline rate generally decreases with time but increases with increasing operating flux or pressure (Table 3), and the specific flux almost always increases with decreasing transmembrane pressure (TMP). For example, an initial specific flux of $733 \text{ L m}^{-2} \text{ h}^{-1} \text{ bar}^{-1}$ has been recorded for a low-pressure submerged P+F system, compared to an equivalent value of $300 \text{ L m}^{-2} \text{ h}^{-1} \text{ bar}^{-1}$ for a side-stream HF system [11,12]. This underlines one of the key aspects of the two configurations. In terms of operational efficiency based solely on liquid pumping, the submerged MBR tends to be much more efficient than the sidestream configuration. This is further discussed in Section 4.

Table 3. Specific flux decline rate for selected MBRs

Membrane configuration	Flux/flux change ($\text{m}^3 \text{ m}^{-2} \text{ d}^{-1}$)	Pressure/pressure change (bar)	Time (d)	Specific flux decline rate ($\text{m}^3 \text{ d}^{-2} \text{ bar}^{-1}$)	Reference
Submerged, P+F, PS 0.4 μm	0.5	0.06	0.04	208	[15]
Submerged, P+F, PS, 0.4 μm	0.1	0.7	0.2	0.79	[14]
Side-stream, tubular, ceramic, 0.1 μm	2.9	1.8	22	0.73	[35]
Submerged, P+F, PS 0.4 μm	0.9	0.06	39	0.38	[15]
Submerged, HF, PE, 0.1 μm	0.2	0.7	70	0.004	[8]
Side-stream, P+F, PS, 2000 kDa	0.5	3.9	165	0.0008	[13]

It is generally recognised that fouling of the membranes is most critical at the start of the filtration run, or when the change in pressure or flux is most rapid [13,14,15,16]. The build-up of the fouling layer in the first few hours of start-up is thus critical in determining operability, and it is this facet of membrane operation that forms the basis of the critical flux hypothesis (Section 2.3.1).

2.2.2 Interactions with membrane material

As already stated, the feed composition can affect permeate flux via interactions with the membrane. For example, the binding of protein and bacteria to hydrophobic

polymeric surfaces has been extensively studied, and the deleterious effect of such

Table 4. Pure water fluxes and protein adsorption for three membranes; $\Delta p = 0.1$ bar, crossflow velocity = 0.05 m s^{-1} [21,23].

Membrane type	Pore size (μm)	Contact angle ($^\circ$)	Permeability ($\text{m}^3/\text{Pa s} \times 10^9$)	BSA adsorption (mg m^{-2})		
				0.1%	1%	10%
Hydrophilic Nylon 66	0.52	<15	34.0	~150	~400	~725
Hydrophilic PVDF	0.72	62.5	17.5	~15	~65	~85
Hydrophobic PVDF	0.80	89.1	21.5	~300	~625	~1150

hydrophobicity on permeability is evident from data reported (Table 4). The contribution of electrostatic and short-range forces to fouling by protein has been demonstrated through pH adjustment [17,18].

Available membrane materials comprise ceramic/metallic (inorganic) or polymeric (organic). Ceramic membranes, such as zirconium and titanium oxide composites, show superior hydraulic, thermal and chemical resistance but are more limited in membrane geometry and significantly more expensive than polymeric materials [19,20]. Naturally hydrophobic polymeric membranes, such as polyethylene (PE), polypropylene (PP), polyvinylidene fluoride (PVF), and polysulphone (PS), can be blended with hydrophilic polymers or surface treated to make them hydrophilic.

Several studies have demonstrated the importance of hydrophobicity in UF and MF membrane materials, and have shown hydrophilic membranes to suffer less from flux decline than hydrophobic ones (Table 4) [21,22,23,24]. For example, Futamura *et al.* [24], using submerged hollow fibre (HF) hydrophobic and hydrophilic PE membranes for filtering domestic wastewater, reported the hydrophobic membrane to give a higher transmembrane pressure (TMP) increase, from 0.5 to 10 bar, over the experimental run than the hydrophilic materials which yielded a stable pressure of 1 bar. Fane *et al.* [22] used hydrophilic and hydrophobic PVDF membranes, operated at 1 bar, for filtering bacterial suspensions, and reported results for the flux recovery following a cleaning cycle. These authors recorded a flux recovery of only 50% for a hydrophobic membrane compared to almost 100% recovery for the hydrophilic membrane, indicating a possible hydrophobic interaction between the membrane and biomass as explained above. On the other hand, Mueller and Davis [25] reported similar fouling characteristics when filtering BSA solutions through both hydrophilic and hydrophobic MF membranes,

indicating the rapid formation of a hydrophobic proteinaceous dynamic layer which subsequently determines the flux, regardless of the nature of the membrane substrate.

Studies of membranes operated on protein solutions suggest that improved performance in filtering activated sludge is obtained using a hydrophilic rather than hydrophobic membranes, since protein makes up a significant percentage of organic matter. However, the degree of hydrophobicity also determines the extent of biofilm attachment onto the membranes. This may then lead to increased bacterial and viral rejection by this protective biofilm layer [8], as bacterial extracellular polymers are comprised in the main of polysaccharides and proteins.

2.3 Fouling remediation

At some point in the filtration process, permeate flux decline reaches a point where it is no longer economical to continue filtration, necessitating membrane cleaning. This is particularly relevant in side-stream operations, where the higher pressures lead to greater fouling. A number of cleaning options exist, with backwashing and/or chemical cleaning being the most common. These all incur additional operating costs from chemical usage and downtime, such that extending the operating cycle by either suppressing fouling or ameliorating problems imposed by it is of key concern in MBR operation.

2.3.1 Sub-critical flux

The ‘sub-critical’ flux hypothesis is that on start-up there exists a flux below which a decline of flux with time does not occur, whereas fouling takes place above this critical value [26]. By correctly selecting the initial flux or TMP the rate of fouling is greatly reduced. The critical value of this flux (or TMP) is very much system specific. For example, Field *et al.* [27], using a flat sheet zirconia-coated metal alloy mesh membrane for filtering incubated yeast cell debris, reported that a TMP of 0.09 bar maintained a flux of $25 \text{ l m}^{-2} \text{ h}^{-1}$ from start-up at a crossflow velocity of 0.74 m s^{-1} . However, once the pressure was increased to 2 bar, there was no proportionate increase in flux, nor was the new flux value of $46 \text{ l m}^{-2} \text{ h}^{-1}$, maintained. The system hydrodynamics is also

important in determining the critical fouling point, as has been demonstrated by investigation of the effect of HF packing density [28]. Generally, there appears to be much corroboratory evidence from MBRs to support the critical flux hypothesis; many submerged P+F plants run at <0.4 bar TMP with no noticeable steady-state flux decline.

2.3.2 *Turbulent aeration*

As with conventional membrane systems, fouling is reduced by operation under turbulent conditions. Turbulent aeration conditions within an MBR promote scouring of the membrane surface to suppress fouling layer formation and, therefore, flux decline [14]. Ueda *et al.* [29], reporting studies based on submerged HF PE membranes, found the airflow rate to determine the cake thickness to some extent, but that there was an optimum value (700 L min⁻¹) above which no further diminution in the cake layer was noted. Airflow rate effects were manifested in the TMP of the system: when air flow was reduced from 700 to 350 L min⁻¹ the TMP increased from 0.2 to 0.5 bar in 20 hours due to the increased solids accumulation on the membrane.

2.3.3 *Cleaning regimes*

The cleaning requirements for the two configurations differ markedly (Table 5). In submerged MBRs the permeate flux is relatively low and can be maintained for extended periods without decline. For example, a submerged plate and frame (P&F) configuration employed a twice-yearly membrane clean with a 0.1% sodium hypochlorite solution [30]. In a side-stream system, the flux and concomitant fouling rate is much higher. Side-stream systems often require more frequent and rather more aggressive cleaning achieved by backwashing (reversal of permeate flow for a short time at filtration pressures) or chemical cleaning. One such system based on a ceramic membrane required weekly cleaning using 5.25% NaOCl heated to 60–80°C along with concentrated nitric acid, the cleaning cycle lasting about 2 hours [31]. Such cleaning conditions demand a very resilient membrane material.

Table 5 Cleaning regimes for a number of MBR systems

Ref:	[16]	[41]	[11]	[29]	[34]	[31]	[12]	[45]
Membrane configuration	Sub P+F	Sub P+F	Sub HF	Sub HF	Side-stream T Yes	Side-stream T Yes	Side-stream HF Yes	Side-stream HF Yes
Cleaned	No	No	No	Yes	Not cleaned; after day 220 fouling became irreversible	Membrane regenerated weekly	Not cleaned for first 15 days	Not cleaned for first 15–70 days then cleaned twice weekly
Cleaning regime				Not cleaned for first 225 d				
Chemical				Water + surfactant		Water + chlorine bleach + nitric acid		

2.3.4 Membrane replacement

Operating costs are significantly affected by membrane replacement frequency, and therefore many of the savings to be realised are achieved by extending membrane life. Generally the more robust ceramic membranes last longer than the less expensive polymeric materials, but the substantial difference in cost of the membrane materials means that membrane replacement costs may be equivalent for the two membrane systems (around 30% in one instance [32]).

3 Biological Activity

3.1 Carbonaceous removal

It has been reported that the membrane in an MBR contributes approximately 30% to the removal of organic matter, this roughly equating to the insoluble fraction [33], with the soluble fraction being removed via the active biomass. Under optimal conditions this biomass is kept as close to a food-limited environment as possible to encourage endogenous respiration. This uncoupling of metabolism, so that catabolism of substrate continues unhindered while anabolism of biomass is restricted, achieves a reduction in the biomass yield [34,35,36] and hence reduced sludge production. MBRs have the advantage of allowing hydraulic retention time (HRT) and solids retention time (SRT) to be independent of each other. Large macromolecules stay within the aeration tank in contact with the biomass for a longer period than that achieved with a conventional activated sludge plant (ASP). This leads to the growth of specialised bacteria and enhanced breakdown and eventual assimilation/mineralisation of the organic matter.

The principle applies as much to anaerobic as aerobic systems and Harada *et al.* [37] showed that in an anaerobic MBR the reactor mixed liquor contained organic matter

with a molecular weight (MW) as large as 10^6 daltons, whereas permeate organic matter had a MW of less than 1500. Low HRTs may be maintained due to the number of adsorption sites available on the smaller, less dense flocs; the greater the number of sites, the greater the load that can be metabolised and adsorbed at the correspondingly high volumetric throughputs [33,39].

From Tables 7 and 8 it can be seen that organic removal is often greater than 95% even with relatively short HRTs of 4–7.5 hours [14,35,38]. Organic loading rates of MBRs are typically higher than that achieved in trickling filters, sequencing batch reactors and conventional ASP, due to the shorter HRTs. MBRs have lower organic loading rates than both high-rate aeration and complete-mix activated ASPs, as well as BAFs, as the organic loading rate is restricted by the permeate flux (see Table 7).

Hanaki *et al.* [39] showed that increasing organic loads to an ASP produces increased heterotrophic activity, as organic matter removal follows first-order kinetics, and this can be assumed to be the case in MBRs. Organic matter removal in MBRs appears not to be significantly affected by low temperatures, such as those between ~ 5 and 20°C [39,40,41,42], which may be due to the number of heterotrophic bacteria in the biomass remaining constant, albeit with a decreased activity at lower temperatures [40].

Table 7 Organic loading rates of different biological treatment processes

Reactor	Organic loading rate ($\text{kgBOD}_5 \text{ m}^{-3} \text{ d}^{-1}$)	HRT (h)	Percentage removal	Reference
BAF: Downflow	1.5 (COD)	1.3	93	[51]
Upflow	4	–	>93	
Downflow	7.5	–	75	
TF: Low rate	0.08–0.4	–	80–90	[47]
Intermediate	0.24–0.48	–	50–70	
High rate	0.48–0.96	–	65–85	
ASP: Sequencing batch	0.08–0.24	12–50	85–95	[47]
Conventional	0.32–0.64	4–8	85–95	
Complete-mix	0.8–1.92	3–5	85–95	
High-rate aeration	1.6–16	2–4	75–90	
MBR: Sub. P+F (Kubota)	0.39–0.7	7.6	99	[14]
Sub. HF (Tech-Sep)	0.03–0.06	1	98–99	[16]
Sub. P+F	0.005–0.11	8	98	[40]
Sub. HF	1.5 (COD)	0.5	87–95	[12]
SS ceramic	0.18	24	>95	[43]
SS P+F	0.45–1.5 (COD)	8	88–95	

3.2 Nitrogen removal

Nitrification has been shown to be greater in an MBR compared with a conventional ASP [39] due to the longer retention times of the nitrifying bacteria (high sludge age, low food/microorganism ratio) and the smaller floc sizes, allowing slightly greater mass transport of nutrients and oxygen into the floc.

From Table 6 it can be seen that MBRs achieve almost complete nitrification [8,14,35,41,43,44] due to the retention of the slow-growing autotrophic bacteria. At nitrogen loads of between 0.1 and 3.3 kgNH₃ m⁻³ d⁻¹, ammonia removal is greater than 90% [8]. As with conventional processes, nitrification is sensitive to feed water quality determinands and operational parameters such as dissolved oxygen (DO), temperature, organic loads, inorganic and organic compounds, pH and levels of key nutrients such as Ca, Mg, Cu and Mo. It has been reported by Chiemchaisri *et al.* [8], that with an influent total nitrogen concentration of between 20 and 50 mg l⁻¹ N, a DO concentration of 1 mg l⁻¹ supplied continuously gives complete nitrification, whereas a DO concentration of 0.5 mg l⁻¹ significantly inhibits this process. Inefficient oxygen transfer rates at low temperatures can lead to reduced activity of strict aerobic bacteria and a reduction in the number of nitrite and ammonium oxidising bacteria [39]. Increasing the DO concentration has been shown to reverse this effect [8], as the process ceases to be oxygen rate limiting.

Hanaki *et al.* [39] showed that heterotrophs out-compete autotrophs for ammonia. Hence, if DO concentrations are high, leading to increased heterotrophic activity, there is less ammonia available for nitrification. Suwa *et al.* [45] verified that nitrification can be maintained at higher rates with lower DO concentrations (<5 mg l⁻¹).

Table 6 COD and NH₃ removal rates in MBRs treating domestic wastewater

Membrane configuration and pore size (μm)	Influent COD (mg l ⁻¹)	Influent NH ₃ (mg l ⁻¹)	COD removal (%)	NH ₃ removal (%)	Reference
Submerged, PE, HF, 0.3 and 0.1	150–450	–	86–97	–	[8]
Submerged, PS, P+F, 0.4	79	–	92	–	[14]
Side-stream, C, T, 0.1	71	35	98.7	97.1	[35]
Submerged, PS, P+F, 0.4	–	25.7	–	91.1	[41]
Side-stream, PS, P+F, 50,000 daltons	488	–	88–94.5	–	[43]
Side-stream, C, 300,000 daltons	–	19.7	–	>99	[44]

Denitrification, the reduction of nitrate to various gaseous end-products such as molecular nitrogen and dinitrogen oxide, can proceed alongside nitrification if:

- aeration is supplied intermittently [8,43],
- hydrodynamics are such that an anoxic area results [43], or
- high organic loads are added allowing anoxic micro-sites to develop within the flocs [43,45,46].

In aerobic MBRs denitrification can be achieved by the addition of an anaerobic tank prior to the aeration tank, with conventional recycle [14]. This produces an environment conducive to the anaerobic denitrifying bacteria to utilise nitrate, giving an effluent much reduced in this nutrient.

4 Energy Consumption

Energy consumption arises from power requirements for pumping feed water, recycling retentate, permeate suction (occasionally), and aeration [32]. The two configurations of utilise some or all of the above, and substantial differences arise between energy consumption of the two MBR operating systems (see Table 8). For example, submerged systems do not require the retentate recycle and some do not require permeate suction.

Energy usage for recycling retentate is proportional to the headloss across the membrane element and the recycle flowrate. The power consumption (P_c) of the recycle pump is given by:

$$P_c = \frac{\Delta p \cdot Q_R}{\eta}$$

where, Q_R = recycle flow rate ($\text{m}^3 \text{s}^{-1}$)

η = pump efficiency (%).

Taking the pump efficiency as 60% [32] and assuming a centrifugal pump with no frictional loss, the energy consumption (E_c) in kWh per m^3 product is given by:

$$E_c = \frac{P_c}{Q_P} \cdot 3.6$$

where, Q_P = permeate flow rate ($\text{m}^3 \text{s}^{-1}$), i.e. the product of the permeate flux and membrane area.

Table 8 Membrane configuration, operating parameters and energy consumption for a number of MBR systems

	[42]	[41]	[11]	[29]	[34]	[31]	[12]	[45]	[54]
Configuration:									
Process	Sub	Sub	Sub	Sub	SS	SS	SS	SS	SS
Membrane	P&F	P&F	HF	HF	T	T	HF??	HF	T
Material	PS	PE	PE	PE	PS	C	C	PS	Al(OH) ₃
Pore size, μm /MWCO, kD	0.4 μm	0.4 μm	0.1 μm	0.1 μm	50 kD	300 kD	0.1 μm	0.1 μm	~0.5 μm
Surface area (m^2)	0.24	0.96	2	4	2.6	0.08	1.1	0.39	-
TMP (bar)	0.1	0.3	0.13	0.15	5	2	2	2.75	2.25
Permeate flux ($\text{L m}^{-2} \text{h}^{-1}$)	7.9	20.8	8	12	170	175	77	8.3	20
Crossflow velocity (m s^{-1})	0.5	0.3–0.5	ND	ND	1–2	3	1.5–3.5	ND	2.2–3.6
Energy consumption, permeate (kWh/m^3)	-	0.013	0.0055	0.23	0.17	9.9	32	0.045	140
Energy consumption, aeration (kWh/m^3 product)	4.0	0.0091	0.14	70	0.52	2.8	9.1	10	10
Total energy consumption (kWh/m^3 product)	4.0	0.022×10^{-2}	0.14	70	0.69	13	41	10	50

KEY

Sub	Submerged	PS	Polysulphone	ND
	No data given			
SS	Sidestream	PE	Polyethylene	
P&F	Plate and Frame	C	Ceramic	
HF	Hollow fibre	MWCO	Molecular Weight Cut-Off	
T	Tubular	kD	kDaltons	

Aeration costs can be calculated from the Δp of the aeration tank (tank height assumed to be 2 m when data not given) and the airflow rate (assumed from literature figures when data not given) [47]. The two MBR configurations utilise aeration in significantly different ways. In the side-stream configuration aeration is supplied to the bioreactor by fine bubble aerators which are highly efficient for supplying oxygen to the biomass (standard oxygen-transfer efficiency (SOTE) of 25–40 % at 5 m depth [47]). In submerged MBRs it is the turbulent aeration which generates the crossflow, at around 1

m s^{-1} (as compared with $2\text{--}4 \text{ m s}^{-1}$ in a side-stream system), as well as scouring the membrane and providing oxygen to the biomass. Coarse bubble aerators are less efficient than fine bubble aerators for supplying oxygen to the biomass (SOTE of 19–37% at 5 m depth) but have the advantage of lower cost [47]. In some operations, e.g. treatment of high-strength liquors, coarse and fine bubble aerators are used in combination [48].

The energy consumption for side-stream systems is usually 2 orders of magnitude higher than that of submerged systems (Table 8). Aeration is the main cost associated with submerged units at >90%, whereas in side-stream operations it is only ~20% of the total cost (Fig. 2). Table 8 indicates that the MF membrane system of Ueda *et al.* [46] demonstrate a higher aeration cost than other submerged systems, but the membrane cleaning requirements reduced as a result (Table 5). One anomaly in Table 8 can be seen in that the side-stream PS HF system [43] has a low permeate flux which may be due to the excessive fouling of the HF membranes by the high level of MLSS.

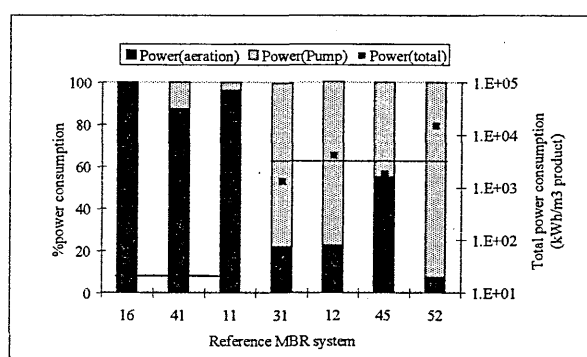


Figure 2 Percentage aeration and permeate energy costs with total power consumption

Figure 2 depicts costs associated with aeration and pumping. It can be seen that the submerged systems have no associated liquid pumping costs unless the permeate is extracted by suction, in which case this can contribute to up to 28% of the costs. The majority of the costs arise from aeration, and the total energy consumption tends to be low in submerged systems yielding high energy efficiencies. However, this is associated with a low flux which increases the membrane requirement and so the capital expenditure. The practical effective size of such systems thus tends to be limited to

5000 population equivalent (p.e.) [30]. In the side-stream systems, it is usually the pumping of the retentate recycle stream that incurs the greatest energy cost, contributing 60–80% of the total costs. The side-stream systems incur much higher operating costs than the submerged systems because of this (Fig. 2).

Packing density and energy consumption are linked: with low packing densities, a higher specific area of membrane is required to produce the same flux and, therefore, the energy requirements increase. The membrane packing density values for the submerged systems, be it P+F or HF, are all very similar and produce similar permeate fluxes (Table 8). It can be seen from side-stream tubular operational data that ceramic membranes demand a smaller surface area and smaller TMPs to produce a similar permeate flux compared with polymeric membranes due to the higher resistance to fouling [19]. The HF system has a lower flux than the tubular system due to its higher fouling propensity.

5 Commercial MBRs

Table 9 outlines commercial MBRs currently available for treating domestic wastewater; there is a more or less equal split between submerged and side-stream configurations. All of the systems run at MLSS levels of up to 20,000 mg/l except for the Rhodia/Orelis system that runs at a lower level of around 13,000 mg/l. The cleaning regimes of the various plants vary from twice yearly chemically cleaning combined with mechanical cleaning every 2–5 years [30], to chemical cleaning every 45 days [49]. One commercially operated submerged MBR, based on the Kubota submerged P&F system, is outlined in Table 10.

Table 9 Commercial MBR plants available for treating domestic wastewater [adapted from 53].

Company	MBR process	Membrane configuration	No. of plants treating domestic wastewater	No. of plants treating industrial wastewater	Capacity ($\text{m}^3 \text{d}^{-1}$)	Type of waste
Rhodia/Orelis, France [49]	Pleiade	Submerged and side-stream; UF; P+F; Acrylonitrile copolymers; 40,000d cut-off	56	14	<7–500	Urban, grey water, black water
Kubota, Japan [50]	Kubota	Submerged; MF; P+F; PS; 0.4 μm	92	50	10–2000	Municipal, grey water, domestic
Zenon, Canada [55]	Zeeweed and ZenoGem	Submerged; MF; HF; polymeric; 0.1 μm	150 total		340–5500	Municipal and industrial
AquaTech, Korea [56]	Biosuf	Side-stream; UF; T; Polyethersulphone; 40,000d cut-off	12	6	40–3000	Municipal
Degrémont, France [57]	MBR	Side-stream	2	3		Municipal

Table 10 Full-scale submerged MBR operational characteristics, Porlock, UK

Parameter	Operational characteristics
Location	Porlock, UK
Function	TSS, COD/BOD and bacteria removal, denitrification
Capacity	1900 $\text{m}^3 \text{day}^{-1}$
Capital cost (million £):	
Sewerage connection	1.5
Ancillary equipment	1.25
Building	1.25
Population equivalent	4000 max
Membrane:	
Configuration	Plate and frame
Material	Polymeric coating on fibrous support
Pore size	0.4 μm
Total area	15,000 m^2
No. units (panels)	150
Cost per 1m^2 panel	£80
Guaranteed life	7 years
Cleaning requirement:	
Chemical (NaOCl)	1/year
Mechanical (jet wash)	1 biannually
Plant footprint (m)	100×20
Influent	Screened raw sewage
BOD (mg L^{-1})	160
Effluent:	
BOD (mg L^{-1})	<4
COD (mg L^{-1})	25
NH_3 (mg L^{-1})	<1
Turbidity (NTU)	<0.3
Faecal streptococci (cfu)	<10

6 Conclusions

- Long SRTs lead to endogenous respiration within the biomass and therefore reduced sludge production. Long SRTs also lead to the growth of specialised bacteria, e.g.

nitrifiers, and enhanced breakdown of large macromolecules. Organic loading rates are restricted by the flux but are generally higher than conventional ASP with greater carbon and nitrogen removal rates.

- Significant bacteria and virus rejection is accomplished with MF and UF membranes. Rejection with MF membranes is enhanced by the build-up of a dynamic layer as well as high turbidity. Disinfection is important not just for EC discharge consents but for water reuse and recycling.
- Rejected matter accumulates at the membrane surface as a concentration polarisation layer that leads to a resistance to filtration and membrane fouling. Flux decline has shown to be important in the early stages of filtration (especially the first 24 hours), and the importance decreases with time.
- Often submerged systems, run at low TMP, achieve a steady-state flux that does not decline over time after the initial fouling period; this is hypothesised to be due to sub-critical flux. Turbulent aeration also reduces the fouling layer build-up by reducing the concentration polarisation layer by scouring the membrane surface, achieving a pseudo-crossflow effect
- Membrane material needs to be hydrophilic to some degree to reduce fouling, especially by proteins and bacteria. Hydrophobic materials lead to increases in TMP or decreases in flux, which in turn lead to increases in energy and cleaning costs to overcome membrane fouling.
- Aeration costs are high in submerged MBRs due to the need to scour the membranes, set up the crossflow effect and to provide oxygen to the biomass. This often leads to coarse and fine bubble aerators being required, compared with only fine bubble aerators in the side-stream system. Aeration in submerged systems is >90% of the total costs whereas only ~20% of the costs derive from aeration in the side-stream systems.

- Submerged systems have very little liquid pumping costs unless permeate flux is enhanced with a suction pump, which can contribute up to 28% of the total costs. However, the low costs are associated with low fluxes compared with side-stream systems, which in turn increases capital costs and footprints.
- Side-stream systems have liquid pumping costs of 60–80% of the total costs, and this along with high cleaning requirements leads to much higher operating costs compared with submerged systems, although the capital costs and footprints are smaller, as the membrane area required is smaller.

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CHAPTER THREE

MEMBRANE BIOREACTORS FOR USE IN SMALL WASTEWATER TREATMENT PLANTS: MEMBRANE MATERIALS AND EFFLUENT QUALITY

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MEMBRANE BIOREACTORS FOR USE IN SMALL WASTEWATER TREATMENT PLANTS: MEMBRANE MATERIALS AND EFFLUENT QUALITY

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Abstract The relative efficiencies of three membrane materials for use in a submerged membrane bioreactor treating domestic wastewater were evaluated. A glass-filled hydrophobic polypropylene membrane allowed no flow through the membrane. A polysulphone membrane (PS, 0.4 μm pore size) did not achieve a steady-state flux although a flux of 7 $\text{l m}^{-2} \text{h}^{-1}$ occurred on the last day of the steady-state run. A melt-blown polypropylene membrane (NWPP, 5 μm pore size) gave a steady-state flux of 5 $\text{l m}^{-2} \text{h}^{-1}$. Membrane productivity was low due to irreversible fouling in the NWPP membrane and reversible fouling in the PS membrane. It was found that the NWPP membrane gave a 5-log reduction in total coliforms whereas the PS membrane gave a 9-log reduction. BOD_5 , COD, SS, $\text{NH}_3\text{-N}$ and turbidity removal was similar with both the PS and the NWPP membranes. It was concluded that although the NWPP membrane was relatively inexpensive and achieved a high degree of organic matter removal, due to its inability to remove bacteria from the effluent it would be inappropriate to apply it to domestic wastewater in a submerged MBR.

Keywords Disinfection; membrane fouling; membrane material; membrane resistance; organics removal; permeate flux

1.0 Introduction

EC legislation regarding effluent discharges from small population equivalent (p.e.) agglomerations demands a treatment process providing an effluent much reduced in, amongst other constituents, BOD/COD, SS, NH_3 and pathogenic bacteria [1,2] One of

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the most promising newer technologies for providing a high quality effluent is the membrane bioreactor (MBR). This is a combined biological and physical process technology for solid/liquid filtration, integrating suspended growth reactors with membrane filtration. The driving force to accomplish separation is the pressure gradient (Δp) across the membrane. The advantage of MBRs over traditional suspended growth reactors is that high concentrations of MLSS, typically up to 20,000 mg/l, can be treated leading to small-scale/high treatment systems with reduced sludge production from longer SRTs. The effluent quality is high due to the retention of colloidal and macromolecular material, including bacteria.

Two configurations of MBR exist, submerged and side-stream. A submerged system operates at a lower transmembrane pressure (TMP) and therefore a lower flux than the side-stream system, but has the advantages of reduced fouling and less rigorous cleaning procedures.[3,4,5] Membrane productivity is expressed as permeate flux through the membrane and this is reduced due to fouling which is caused due to a build-up of retained material on or inside the membrane and which may be partially recovered by mechanical or chemical cleaning. One of the negative aspects of membrane filtration for domestic wastewater treatment is the high cost of the membrane itself; high areas are required to meet the design flow of the plant.

This paper presents the results from a pilot plant study using three different membranes for use in a submerged MBR treating domestic wastewater. The ability of the microfiltration membranes to remove organics and pathogenic bacteria are investigated with respect to the current legislation in the EU. Membrane fouling is also investigated with regard to reversible and irreversible fouling propensity.

2.0 Materials and Methods

A schematic of the pilot plant is shown in Figure 1. Influent was pumped from the main feed tank (A) to a constant head device (5 l) (B) that fed the bioreactor, equalising feed and permeate flow. Each bioreactor was made of perspex with a working volume of 0.035 m³. The feed flow to the constant head device was higher ($\sim 0.5 \text{ l min}^{-1}$) than that required to feed the bioreactor ($\sim 0.35 \text{ l min}^{-1}$) so as to keep the influent fresh and mixed;

the surplus overflowed to the waste tank (C). The feed pressure to the bioreactor was set at 0.06 bar (60 cm water) by the constant head device. Two plate and frame membranes were immersed in each bioreactor, held upright by a perspex sleeve, giving a membrane surface area of 0.24 m^2 (D). Compressed air was supplied through aerator stones (F) via a compressor and the airflow rate was regulated by an airflow metre. Air was supplied at 20 l min^{-1} (G) for the first 38 days to allow the system to reach steady-state. On day 39 the airflow rate was reduced to 5 l min^{-1} for 13 days to investigate the effect on permeate flux and fouling. On day 52 the airflow was increased back to 20 l min^{-1} . At times a suction pump was connected to the permeate side to increase the TMP. Permeate samples were collected from the sample tubes (E). Waste drains fitted at the base of the reactor were used to empty the bioreactor when required (H).

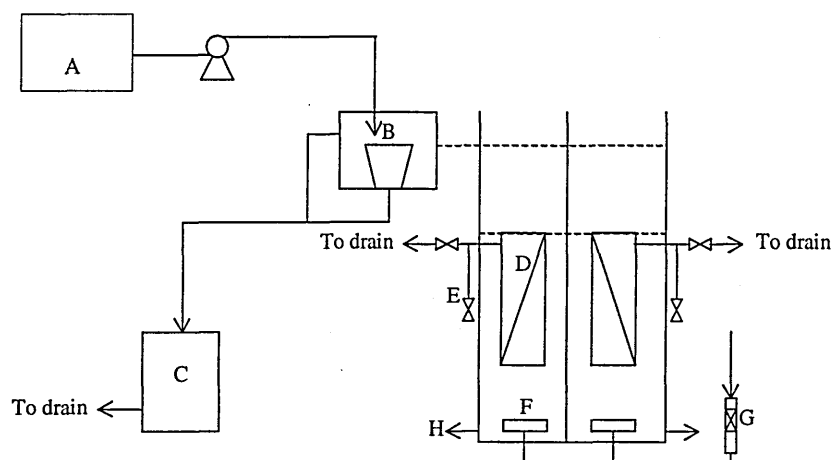


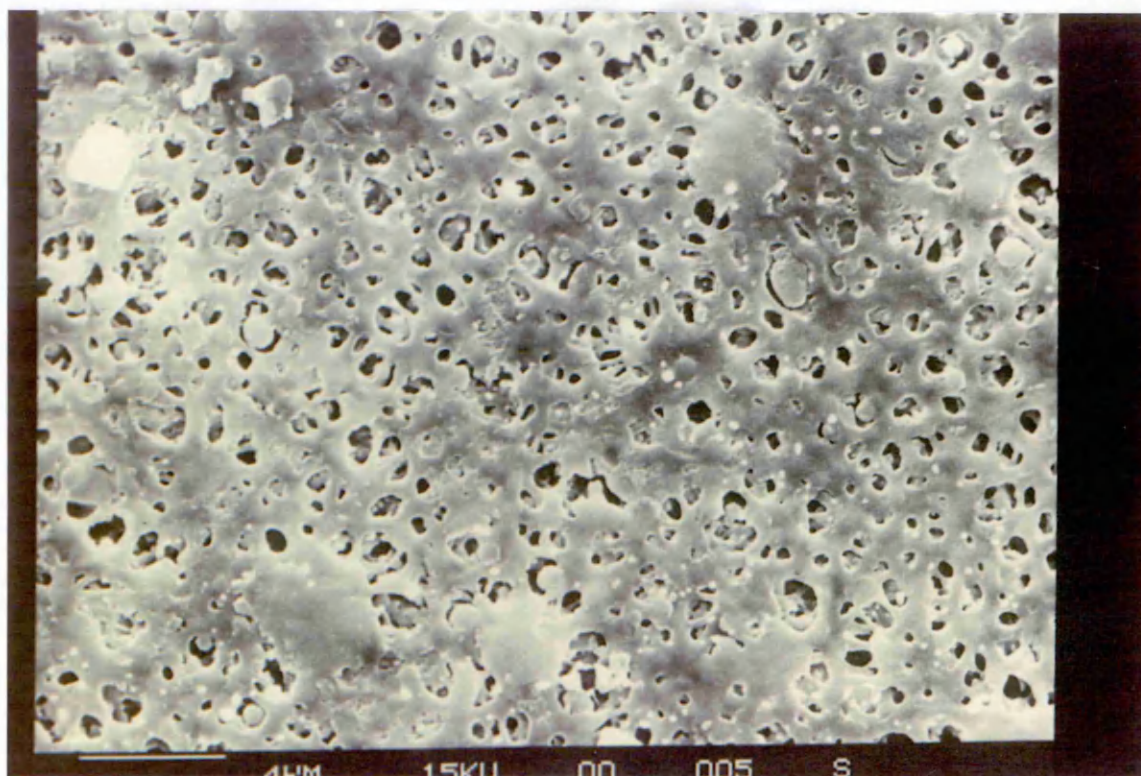
Figure 1 Schematic of the pilot plant MBR

The flat plate PVC membrane module had four drainage channels on the permeate side with a cross-hatched area in-between for maximum permeate flow. The membranes were attached with silicone sealant and consisted of a polymeric upper layer and a fibrous support layer. A number of membranes were tested (Table 1) and their descriptions follow below with scanning electron micrographs (SEMs) of their upper surface (Figures 2–4).

From the SEMs it can be seen that the different materials have different surface characteristics (Figures 2–4). For example, the polysulphone (Figure 2) shows a smooth surface with pores distributed over the surface. The glass-fibrilated material shows the number of small glass beads embedded in the material which have been used for the production of the pores (Figure 3) and the non-woven material is highly tortuous (Figure 4). Both the latter materials are obviously homogeneous, unlike the classical anisotropic membrane shown in Figure 2. They are also mechanically unstable, readily stretchable in both dimensions.

Table 1 Membrane materials tested

Material	Mean pore size (μm)	Contact angle ($^{\circ}$)	Clean water flux ($\text{L m}^{-2} \text{h}^{-1}$)
Polysulphone	0.3	~52	175
Glass-fibrilated polypropylene	0.4	~16	0
Non-woven polypropylene	5	~40	232.5

Figure 2 Upper surface view of 0.4 μm polysulphone membrane material

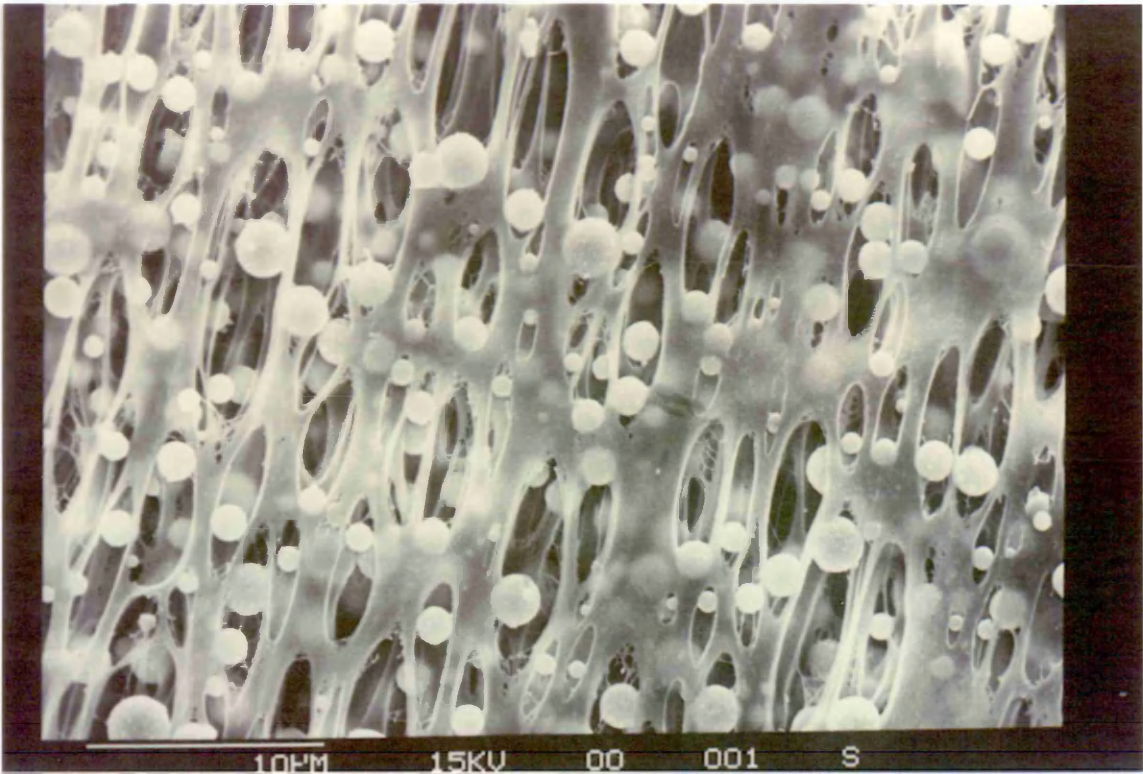


Figure 3 Upper surface view of 0.5 μm glass-fibrilated polypropylene membrane material

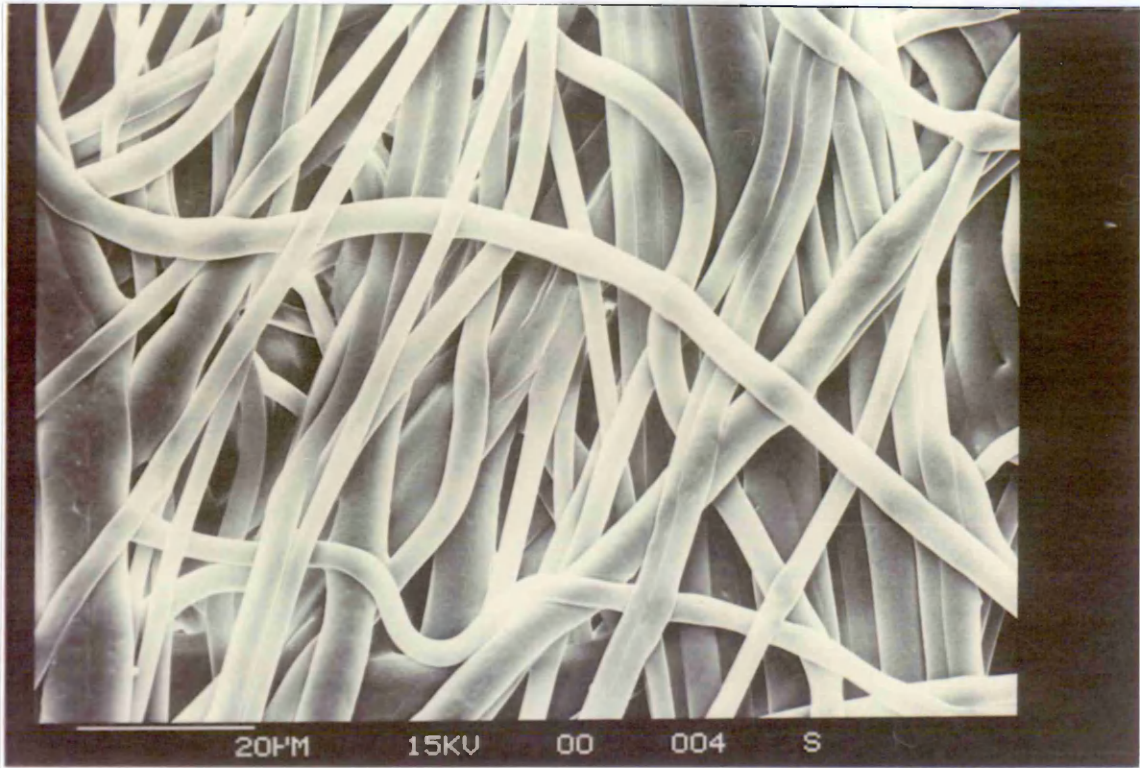


Figure 4 Upper surface view of 5 μm non-woven polypropylene membrane material

The membranes were cleaned mechanically with tap water and a sponge on day 56. The feed supplied to the system was screened, settled, primary influent and was supplied via a constant head device such that permeate flow was equal to feed flow. The characteristics of the wastewater are given in Table 2.

Table 2 Influent characteristics

Constituent	Value
BOD ₅ (mg l ⁻¹)	134
Total COD (mg l ⁻¹)	250
NH ₃ -N (mg l ⁻¹)	16.5
SS (mg l ⁻¹)	78.6
Turbidity (NTU)	79.7
Total coliforms (cfu 100 ml ⁻¹)	26.7×10 ⁶

Prior to start-up, clean water fluxes were carried out *in-situ*, using tap water recirculated back to the bioreactor from the permeate drains. Clean water measurements were also carried out *ex-situ* using a vertical hydrostatic head at 0.06 bar but under orthogonal flow conditions, with no mixing. The fouled membrane was more rigorously cleaned than in the previous experiment, as may be applicable in practice. Mechanical and chemical cleaning was carried out with a sponge and surfactant for 2 min under running water.

BOD₅ and SS analyses were carried out in accordance with Standard Methods (1992, Methods 5210 A and 2540 D, respectively).[6] COD analysis was carried out in accordance with a US EPA approved method utilising Hach Laboratory Method 8000 (Spectrophotometer Model DR/2010). NH₃-N analysis was carried out using Hach Laboratory Method 10031 (Spectrophotometer Model DR/2010). Turbidity was analysed using a Turbidimeter Model 2100N (Hach Laboratory). Total coliforms were analysed using the IDEXX Quanti-Tray 2000.

3.0 Results and Discussion

3.1 Glass-filled polypropylene membrane

The first membrane used, the hydrophobic GFPP with a 0.4 μm pore size, did not allow a flow through the membrane. This was attributed to its degree of hydrophobicity and the MBRs low operating pressure.

3.2 Polysulphone and non-woven polypropylene membranes

3.2.1 Flux decline

Figure 5 shows the permeate fluxes of the two membranes. It can be seen that the initial flux from the NWPP membrane was high at $162.5 \text{ l m}^{-2} \text{ h}^{-1}$. This rapidly declined to 11.25 in 2 hours and to $8 \text{ l m}^{-2} \text{ h}^{-1}$ in 3 hours. A steady-state flux of $\sim 5 \text{ l m}^{-2} \text{ h}^{-1}$ was reached. The PS membrane had an initial flux of $45 \text{ l m}^{-2} \text{ h}^{-1}$, which declined to $41.75 \text{ l m}^{-2} \text{ h}^{-1}$ after 3 hours. No steady-state flux was achieved but on day 39 the flux had reduced to $7 \text{ l m}^{-2} \text{ h}^{-1}$. It could be argued that the steady-state flux was operating below the critical flux value and, therefore, a decline in flux would not occur with time.[7] This may be important in operating costs as reduced fouling leads to reduced cleaning requirements and an increased longevity of the membrane.

On day 39 the airflow rate was reduced to 5 l min^{-1} and there was no decline in flux. This showed that the bubbles produced from the diffuser were still turbulent enough to scour the membrane surface and reduce the fouling layer.

Membrane cleaning, which was carried out on day 57, significantly enhanced permeate flux from both the membranes. The NWPP membrane showed a flux of 52.50 and the PS membrane a flux of $29.75 \text{ l m}^{-2} \text{ h}^{-1}$. These figures show that after cleaning, the PS membrane demonstrated almost complete flux recovery, whereas the NWPP membrane showed relatively little flux recovery with reference to the virgin material flux of $162.5 \text{ l m}^{-2} \text{ h}^{-1}$. This would seem to indicate that either pore size or pore structure (or some macroscopic property reflecting pore structure) is instrumental in determining the extent of irreversible fouling, since both membranes are ostensibly hydrophilic

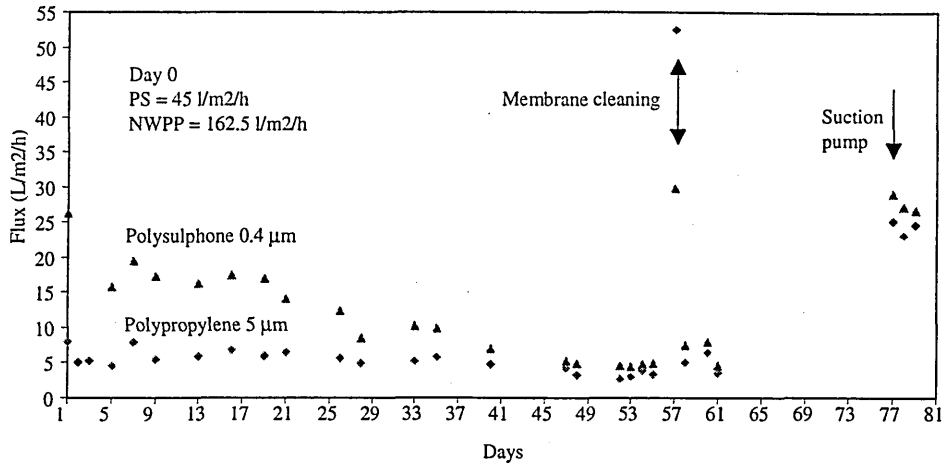


Figure 5 Flux decline of PS and NWPP membranes over period of experimental run. Note the increases in permeate flux when membrane washing and addition of a suction pump occurred.

On day 77 a suction pump (TMP of 0.2 bar) was attached to the effluent stream, to investigate the effect on permeate flux. From Figure 5 it can be seen that the flux increased from 4.5 to 29 l m⁻² h⁻¹ from the PS membrane and from 3.5 to 25 l m⁻² h⁻¹ from the NWPP membrane. The consequences of this, apart from an increase in permeate flux, was an initial increase in COD and in total coliforms. After two days both COD and coliform counts fell back to their previous levels in both cases.

Clean water permeability studies allow the hydraulic resistance of the membrane to be measured, while permeabilities attained after the fouled membranes have been mechanically cleaned reflect the degree of reversible fouling (Table 3). The resistance of the fouling layer according to the simple resistance model is:

$$J = \frac{\Delta p}{\mu R_t} = \frac{\Delta p}{\mu(R_m + R_f)}$$

where Δp = transmembrane pressure (kg m⁻¹ s⁻²)

μ = viscosity (assumed to be 10⁻³ throughout, kg m⁻¹ s⁻¹)

R_t = total hydraulic resistance (m⁻¹)

R_m = clean water membrane resistance (m⁻¹)

R_f = partially cleaned membrane resistance (m⁻¹)

Table 3 Membrane and fouling layer resistances

Membrane	Clean water resistances		MBR resistances		Membrane ratios			
	R_m	R_f	R_m'	R_f'	R_m/R_f	R_m'/R_f'	R_m/R_m'	R_f/R_f'
PS	4.4×10^{10}	1.4×10^{11}	4.8×10^{11}	7.2×10^{11}	0.31	0.67	0.09	0.19
PP	3.6×10^9	3.0×10^{11}	1.3×10^{10}	4.1×10^{11}	0.01	0.32	0.03	0.73

The R_m/R_f ratio demonstrates that flux from the PS membrane is more readily recoverable compared with the NWPP membrane as the fouling layer of this membrane offers more hydraulic resistance even when mechanically and chemically cleaned. The PS R_m'/R_f' ratio demonstrates that the membrane has a greater recovery due to the cleaning procedure relative to the initial operation in the actual MBR compared with the NWPP membrane. Both of these ratios indicate that fouling of the NWPP membrane may be internal, since mechanical and mild chemical cleaning appears to result in good flux recovery in either use.

The R_m/R_m' ratio shows the effect of the start-up procedure on membrane fouling. This procedure requires the activated sludge inoculum to be held within the aerated bioreactor for 24 hrs before the permeate drains are opened. It appears that the NWPP membrane exhibits greater initial fouling compared with the PS membrane, which may be a function of internal fouling occurring in the 24 hrs.

The R_f/R_f' ratio shows the effect of cleaning on the fouled membranes. It can be seen that the more rigorous cleaning regime employing surfactant is especially effective on the PS compared with the NWPP membrane. This may demonstrate the ability of the surfactant to help dislodge the surface gel-layer on the PS membrane, whilst still being unable to penetrate the NWPP membrane to remove internal fouling.

3.2.2 Total coliform removal

Figure 6 shows the log reduction of total coliforms from the two membranes. Effluent from the PS membrane showed a 9-log reduction for total coliforms, whereas the NWPP membrane showed a 5-log reduction. It can be seen that influent concentration shows a

natural variation. Both of the membranes show their respective log reduction regardless of this natural. When the suction pump was attached on day 69, which produced a doubling of the flux, both the membranes showed coliform breakthrough for 2 days before the membranes adapted to the increased hydraulic load and recovered.

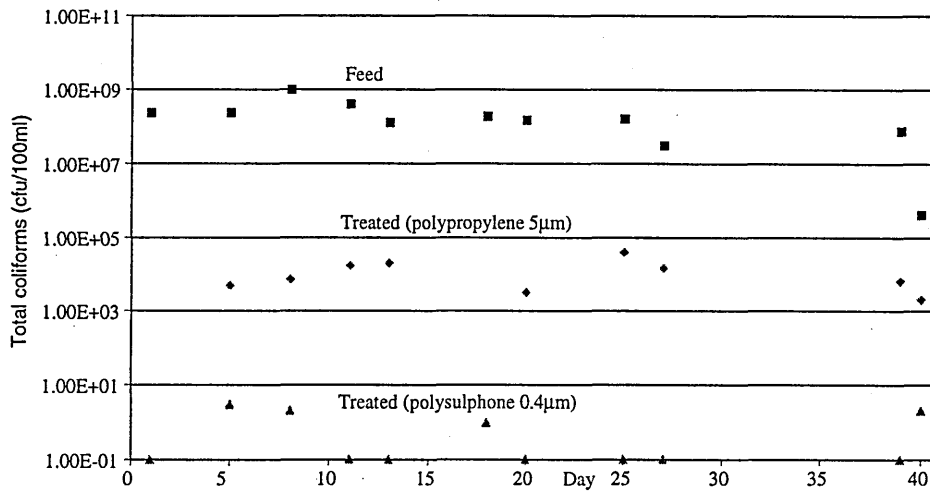


Figure 6 Log removal of total coliforms from PS and NWPP membranes

The PS membrane complies with both the mandatory and guide limit for the total coliform requirement in the Bathing Water Directive (see Table 5), whereas the NWPP does not meet either of these limits.[2] Coliform removal is also an important quality criteria for water reuse. For example, the Building Services Research and Information Association and the EPA (Florida) sets a standard of 0 faecal coliforms per 100 ml for water that is to be reused, e.g. for WC flushing.[8] This is the main difference between the two membranes: the PS membrane disinfects and the NWPP membrane does not, and the latter therefore does not meet the Bathing Water Directive or the water reuse standards.

3.2.3 Organics removal

For the two bioreactors the mean organic loading was $0.2695 \text{ kg BOD m}^{-3} \text{ day}^{-1}$ for the PS membrane and $0.1155 \text{ kg BOD m}^{-3} \text{ day}^{-1}$ for the NWPP membrane. Figure 7 shows that within the organic loading rate the mean BOD_5 removal efficiency remained higher than 97%. The NWPP membrane had a wider range of organic loading which was

almost completely removed compared with the PS membrane, due to the higher flux, even when the MLSS fell to $<5000 \text{ mg l}^{-1}$. This ability to absorb natural variations in feedwater quality loading rates indicates that the MBR process is very robust.

The organic load is removed efficiently even at the low MLSS values due to the extremely low F/M ratios (0.08 and 0.04 for the PS and PP membrane, respectively). The low F/M values arose as the sludge age of the system approached infinity (i.e. no sludge wastage). This has important economic considerations as an MBR operated with no sludge wastage would obviously lead to zero sludge storage or disposal costs. Sludge disposal costs make up a significant percentage of the total annual revenue expenditure of water companies in the UK (13% in 1975/6).[9] Even with these low F/M ratios the ratio of MLSS/MLVSS stays above 98%. This is an important consideration in MBRs since inert materials cannot be degraded and so will always accumulate in the sludge. A consequence of low MLSS combined with good organic removal would be a lower energy demand for mixing and oxygen transfer within the bioreactor. It is the combined mixing, scouring and oxygen transfer requirements from the aeration system, which prove to be a significant cost in the submerged MBR system.

Table 4 Percentage reductions in relation to the influent load of some organic constituents

Constituent (% reduction)	PS	NWPP
BOD ₅	97.4	96.9
Total COD	92.4	93.3
NH ₃ -N	97.6	99.8
Turbidity	99.8	99.6

BOD₅ and COD removal from both membranes was higher than that required by current legislation (see Table5) as was SS. The average permeate turbidity from both the MBRs were very low at $<1 \text{ NTU}$, representing $>99.5\%$ reduction (Table 4). BOD and turbidity are also constituents that are important for water reuse.[9]

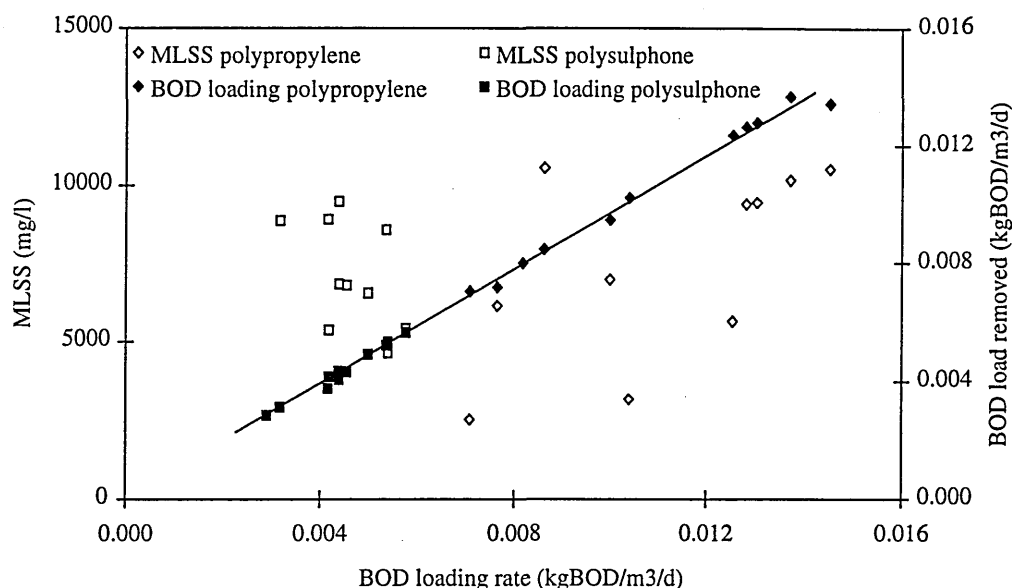


Figure 7 BOD loading rate with respect to load and percentage removal for PS and NWPP membranes

Table 5 EC legislation for effluent discharges from small wastewater treatment plants

Constituent	Percentage reduction	Mandatory limit	Legislation met		Reference
			PS	NWPP	
BOD ₅ (mg l ⁻¹)	70–90	25	Yes	Yes	EEC, 1991
COD (mg l ⁻¹)	75	125	Yes	Yes	EEC, 1991
SS (mg l ⁻¹)	35	35	Yes	Yes	EEC, 1991
Total coliforms (cfu 100 ml ⁻¹)		10,000	Yes	No	EEC, 1976

3.3 Economic Factors

The benefits of operating at low pressures and therefore low flux is a reduced fouling level which leads to a less rigorous cleaning and replacement requirement. The negative aspect is the large area of membrane needed, which increases the capital costs. One particular P+F membrane for use in submerged MBRs has a cost of £50 m⁻² at this time. The alternative NWPP is an ostensibly simpler, homogenous material which would incur lower manufacturing costs, say £15–25 m⁻².

4.0 Conclusions

- The membrane and fouled membrane resistances indicate that the PS membrane showed reversible fouling and the NWPP membrane irreversible fouling.

- Both the NWPP and PS membranes produced an effluent extremely low in organic matter well within the current legislation requirements.
- The NWPP membrane, although removing >95% of total coliforms, did not produce an effluent that meets current legislation or water reuse standards compared with the PS membrane.
- The robustness of MBRs is indicated by their ability to absorb natural variations in load without an adverse effect on treatment capability.
- The ability of MBRs to produce a high quality effluent even when MLSS were low may lead to zero sludge disposal costs.
- Although the NWPP membrane is relatively inexpensive to manufacture compared with the PS membrane, its inability to adequately remove coliforms means that it would be inappropriate for use in submerged MBRs treating domestic wastewater.

5.0 Acknowledgements

The authors wish to thank the EPSRC and Severn Trent Water for supporting this work.

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CHAPTER FOUR

LOW-COST POLYMERIC MEMBRANES FOR MEMBRANE BIOREACTORS

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LOW-COST POLYMERIC MATERIALS FOR MEMBRANE BIOREACTORS

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Abstract The efficacy of low-cost extruded and melt-blown non-woven polypropylene membranes for wastewater treatment by a membrane bioreactor has been evaluated, based on both submerged and sidestream configurations. The sidestream configuration (SidMBR) has been challenged with both real and analogue sewage matrices. The submerged configuration (SubMBR) was operated concurrently with an identical MBR based on a more conventional polysulphone membrane.

Results for both sidestream and submerged configurations indicated severe flux decline due to fouling, even when a rudimentary cleaning cycle was introduced for the SSMBR. However, results from trials on a kaolin analogue indicated that the kaolin provided substantial protection of the SidMBR membrane, with increased flux and bacterial rejection over corresponding conditions for secondary sewage and mains water filtration. Globular protein, on the other hand, appeared to both catastrophically decrease the flux and enhance the transport of bacteria through the membrane. For the submerged configuration only the non-woven material gave a steady-state flux, but this was combined with a low bacterial rejection when compared with the PS membranes. Removal of BOD₅, COD, SS, NH₃-N and turbidity, on the other hand, was similar for both the PS and the NWPP membranes.

Keywords Membrane bioreactors; disinfection; membrane fouling; membrane material

1.0 Introduction

Two configurations of MBR exist, submerged and side-stream. A submerged system operates at a lower transmembrane pressure (TMP) and therefore a lower flux than the side-stream system, but has the advantages of reduced fouling and less rigorous

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cleaning procedures.[1,2] The sidestream configuration operates at higher Reynolds numbers and so offers better control of concentration polarisation than the submerged configuration. The power consumption, on the other hand, tends to be higher due to the increased TMP.

Membrane productivity is expressed as permeate flux through the membrane. Crucial to the viability of membrane bioreactors is the:

- fouling and subsequent cleaning of the membrane, and
- cost of the membrane.

Since the flux is limited by fouling, caused by a build-up of retained material on or within the membrane, suppression of fouling or the amelioration of its effects is crucial. On the other hand, low-flux operation can be viable if both fouling propensity and membrane cost are low enough.

This paper presents the results from a pilot plant study based on ostensibly low-cost polymeric microfiltration membrane materials. Sidestream treatment has been assessed with reference to physical filtration of secondary effluent using a 25-mm diameter tubular extruded polypropylene membrane. The submerged configuration was based on an A4-sized plate and frame module of both extruded polymer and melt-blown non-woven fibrous material, and was challenged with screened sewage. The ability of the microfiltration membranes to remove organics and pathogenic bacteria have been investigated, along with their fouling propensity as reflected in the flux decline.

2.0 Materials and Methods

2.1 Sidestream treatment

The plant consisted of a 2.4 m length of the tubular extruded polypropylene (EPP) membrane, 25 mm dia., operated in crossflow mode at a transmembrane pressure (TMP) of around 1.5 bar. Cleaning was accomplished by periodic collapsing of the membrane, under reduced feed pressure, automated by electronic timers actuating the

feed pump. Cycle lengths of 10 minutes on / 0.5 minutes off were used throughout. Further details of the plant have been provided elsewhere.[3]

Feed solutions comprised primary or secondary effluent from municipal wastewater treatment (Table 1), based on primary sedimentation and secondary trickling filtration, and sewage analogues. The analogues comprised suspended solids (SS) as kaolin (mean particle size $\sim 2\mu\text{m}$) and dissolved protein as bovine serum albumin fraction V (BSA). Analogues were spiked with appropriate quantities of bacteria to produce levels similar to those recorded for sewage.

2.2 Submerged bioreactor

The submerged MBR experiments were run concurrently with a polysulphone-based membrane and the test membranes, of identical plate and frame design. Each bioreactor was made of perspex with a working volume of 0.035 m^3 and contained membranes having a total surface area of 0.24 m^2 . The polysulphone (PS) membrane had a pore size of $0.4\text{ }\mu\text{m}$ and was modified to give a hydrophilic surface. The two low-cost polypropylene membranes tested were the EPP and a non-woven polypropylene (NWPP) with a $5\text{ }\mu\text{m}$ pore size and surface-treated to create an ostensibly hydrophilic surface.

The TMP was supplied by a constant hydrostatic head of 60 cm water. The feed water of screened, settled, primary influent (Table 1) was provided via a constant head device. Compressed air was supplied via aerator stones at the bottom of the reactor at 20 l min^{-1} for the first 38 days to allow the system to reach steady-state. The airflow rate was reduced to 5 l min^{-1} on day 39 for 13 days to investigate the effect on permeate flux and fouling, and increased back to 20 l min^{-1} on day 52. The membranes were cleaned mechanically with tap water and a sponge on day 56.

BOD₅ and SS analyses were carried out in accordance with Standard Methods (1992, Methods 5210 A and 2540 D, respectively). COD analysis was carried out in accordance with a US EPA approved method utilising Hach Laboratory Method 8000 (Spectrophotometer Model DR/2010). NH₃-N analysis was carried out using Hach

Laboratory Method 10031 (Spectrophotometer Model DR/2010). Turbidity was analysed using a Turbidimeter Model 2100N (Hach Laboratory). Total coliforms (TC) were analysed using the IDEXX Quanti-Tray 2000.[4]

Table 1 Municipal sewage effluent quality determinants, mean values

Parameter	Primary Effluent	Secondary Effluent
BOD ₅ (mg l ⁻¹)	134	-
Total COD (mg l ⁻¹)	250	142
Protein (mg l ⁻¹)	87.3	81.5
SS (mg l ⁻¹)	80	113
Turbidity (NTU)	80	78.0
Total coliforms (TC)	2.6×10^7	3.9×10^6

3.0 Results and Discussion

Sidestream flux data are normalised with respect to TMP, giving flux in lmh per bar (which is then directly proportional to permeability), and mainly refer to the 0.45 μm extruded PP membrane.

3.1 Sidestream treatment

3.1.1 Membrane collapsing and the operational cycle

The effect of membrane collapsing and the cyclic operation itself is demonstrated in Figure 1 with respect to (a) the overall filter run and (b) an individual cycle.

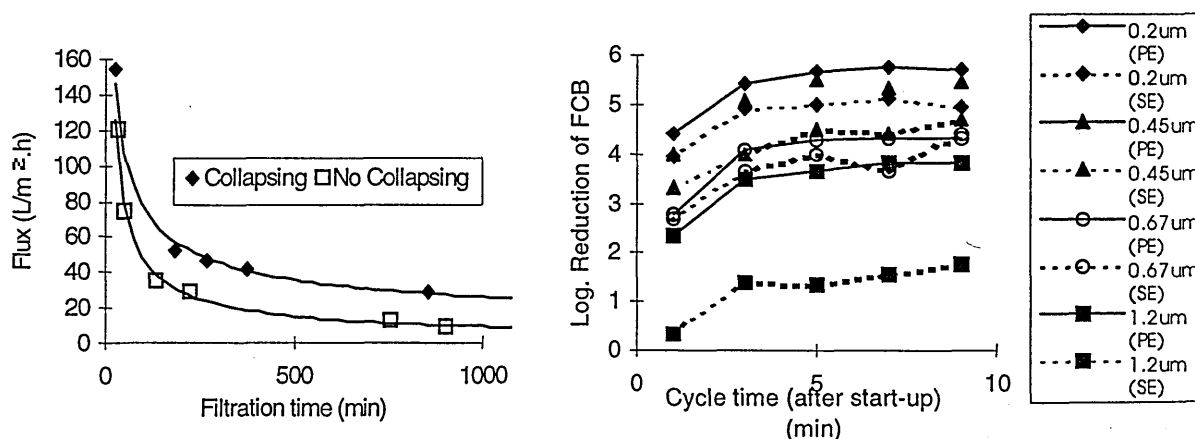


Figure 1 Effect of membrane collapsing: (a) flux over the course of the filter run (SE, 3.8 m/s crossflow), and (b) rejection of faecal coliform bacteria over the course of a single operational cycle (PE = primary effluent; SE = secondary effluent)

Whilst cyclic collapsing of the membrane yields a marked improvement in the flux (more than doubling the equilibrium value) the rejection decreases at the start of the cycle as a result of the pulsing of the pressure. All rejection data reported refers to the mean rejection taken over the complete cycle; rejection was found to be largely independent of filtration time after a filtration time of ten minutes-or-so. Reported flux data all refer to a filtration time of 240 minutes, normally representing approximately twice the equilibrium value in cases where fouling was taking place.

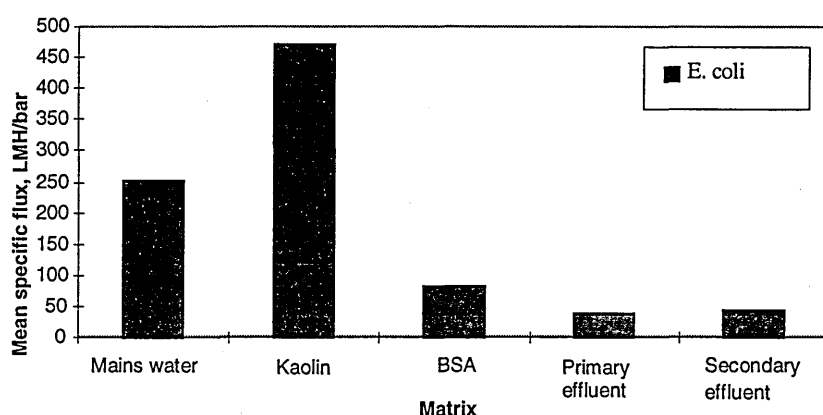


Figure 2 summary of comparative specific flux values for all matrices after 240 minutes, 0.4 μm EPP membrane, averaged over crossflow velocity of 0.9-5.7 m/s. The flux is around double the equilibrium value for sewage effluents

3.1.2 Fouling and rejection

Figure 1 shows flux decline to be significant, even when the cleaning cycle is applied. It appears that the membrane is prone to internal fouling - even when challenged with mains water. The latter point is illustrated in Figure 2, which compares the 240-minute specific flux values for mains water, primary effluent (i.e. settled sewage), secondary effluent (from biological treatment) and the two analogues of kaolin and BSA. Unsurprisingly, the flux decline when the membrane is challenged with BSA is severe. The kaolin, however, appears to offer substantial protection of the membrane, such that the flux measured for filtration of the kaolin suspension is actually higher than that recorded for mains water.

Bacterial rejection (Figure 3) indicates that, contrary to expectation, the decreased permeability of the protein-fouled membrane does not give rise to a correspondingly enhanced rejection, but rather the reverse. Protein appears to assist the transport of bacteria, by about an order of magnitude over the mains water value. The kaolin dynamic layer, on the other hand, increases rejection. COD rejection data for the sewage challenge tests were slightly higher than those for protein rejection at 66-82%, cf. 59-74%.

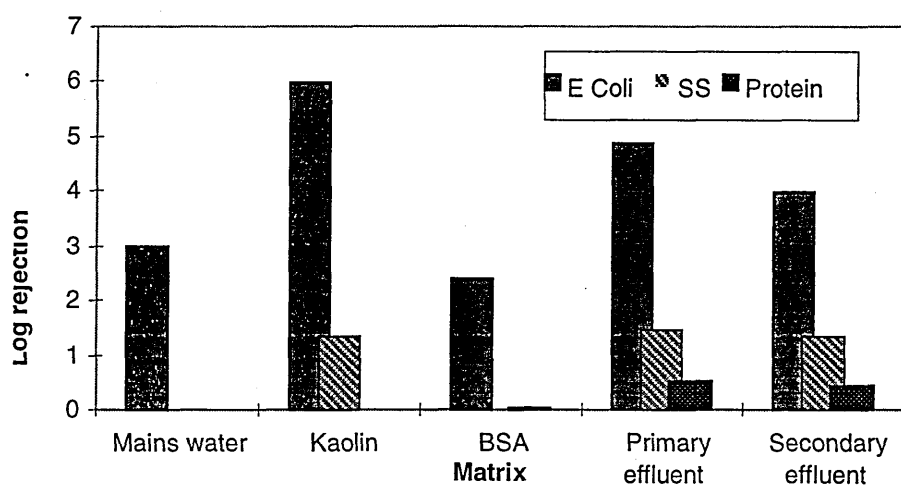


Figure 3 Summary of rejection values for all matrices, 0.45 μm EPP membrane, averaged over crossflow velocities of 0.9-5.7 m/s. Data refer to feed analogue levels of 50-150 ppm kaolin or BSA

3.2 SubMBR, polysulphone, extruded and non-woven polypropylene

The first membrane used, the hydrophobic EPP with a 0.45 μm pore size, did not allow a flow through the membrane. This was attributed to its degree of hydrophobicity and the low TMP employed.

3.2.1 Fouling

Figure 4 shows the permeate flux decline for the PS and NWPP membranes. It can be seen that the initial flux from the NWPP membrane was high at $162.5 \text{ l m}^{-2} \text{ h}^{-1}$. This rapidly declined to $11.25 \text{ l m}^{-2} \text{ h}^{-1}$ in 2 hours and to $5 \text{ l m}^{-2} \text{ h}^{-1}$ in 3 hours where it stabilised, giving a specific flux of $83 \text{ l m}^{-2} \text{ h}^{-1} \text{ bar}^{-1}$. The PS membrane had an initial flux of $45 \text{ l m}^{-2} \text{ h}^{-1}$, which declined to $41.75 \text{ l m}^{-2} \text{ h}^{-1}$ after 3 hours. No steady-state flux was achieved but on day 39 the flux had reduced to $7 \text{ l m}^{-2} \text{ h}^{-1}$, giving a specific flux value of

$117 \text{ l m}^{-2} \text{ h}^{-1} \text{ bar}^{-1}$. The airflow rate was then reduced to 5 l min^{-1} with no subsequent decline in flux, suggesting mass transfer of water through the membrane to be largely independent of the scouring action of the bubbles produced from the diffuser over this range of airflow.

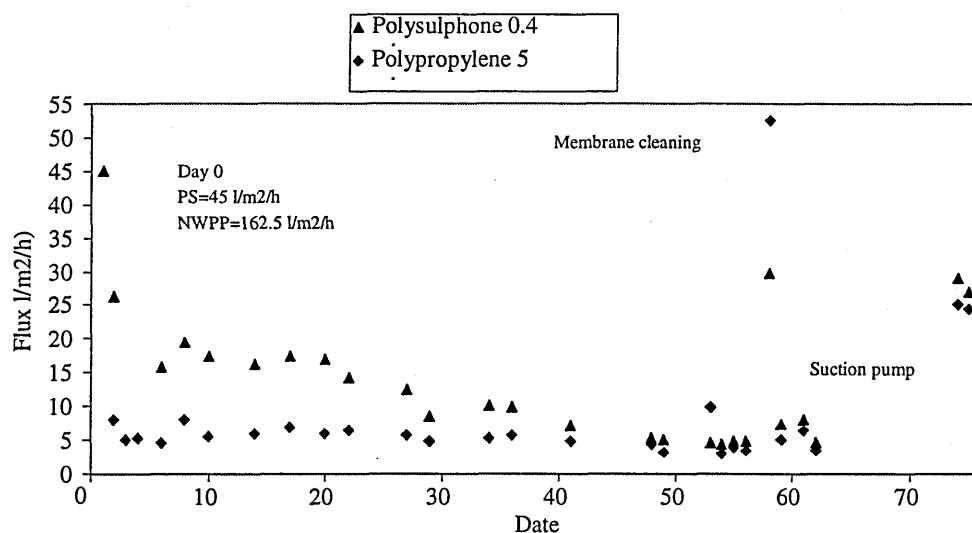


Figure 4 Flux decline of PS and NWPP membranes; the flux increases arise from membrane washing and permeate suction

Membrane sponge cleaning, carried out on day 57, significantly enhanced permeate flux from both the membranes. The NWPP membrane flux increased to $52 \text{ l m}^{-2} \text{ h}^{-1}$ and the PS membrane flux to $30 \text{ l m}^{-2} \text{ h}^{-1}$. These figures show that, following cleaning, the PS membrane demonstrated almost complete flux recovery, whereas the NWPP membrane showed little flux recovery with reference to the virgin material flux of $162.5 \text{ l m}^{-2} \text{ h}^{-1}$. This indicates that either pore size or pore structure (or some macroscopic property reflecting pore structure) is instrumental in determining the extent of irreversible fouling, since both membranes are ostensibly hydrophilic. On day 77 a suction pump (TMP of 0.2 bar) was attached to the effluent stream, to investigate the effect on permeate flux. From Figure 2 it can be seen that the flux increased from 4.5 to $29 \text{ l m}^{-2} \text{ h}^{-1}$ from the PS membrane and from 3.5 to $25 \text{ l m}^{-2} \text{ h}^{-1}$ from the NWPP membrane.

3.2.2 Rejection of microorganisms

Figure 5 shows rejection of total coliforms from the two membranes. Effluent from the PS membrane showed a 9-log reduction for TC, whereas the NWPP membrane yielded a 5-log reduction only after several days of operation. It can be seen that influent concentration shows a natural variation, and rejection by the membranes was maintained regardless of this variation. When the suction pump was attached on day 69, which produced a doubling of the flux, both the membranes suffered coliform and COD breakthrough for 2 days before recovering to previous levels.

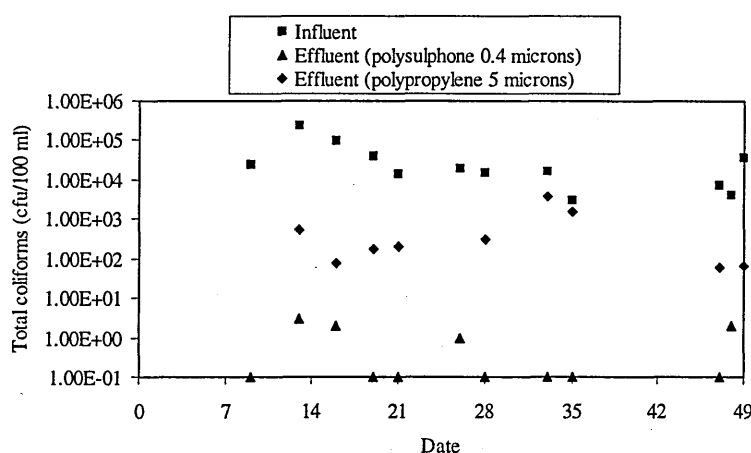


Figure 5 Log removal of total coliforms from PS and NWPP membranes

3.2.3 Removal of organics and ammonia

Results for biodegradable components suggest that removal of these materials is not dependent on the membrane material (Table 2). BOD₅, COD and ammonia were all removed to about the same extent by both MBRs. The longer hydraulic residence time (HRT) for the lower-flux NWPP MBR gives rise to complete nitrification (3-log reduction in ammonia) compared with 2-log reduction for the PS MBR.

For the two bioreactors the mean organic loading was 0.2695 kg BOD m⁻³ day⁻¹ for the PS membrane and 0.1155 kg BOD m⁻³ day⁻¹ for the NWPP membrane. Figure 6 shows that within the organic loading rate the mean CBOD₅ removal efficiency remained at around 97% for both MBRs. Performance was largely independent of feedwater quality: the NWPP achieved almost complete organics removal even when the MLSS fell to

<5000 mg l⁻¹. This arises from the extremely low F/M ratios (0.08 and 0.04 for the PS and PP membrane, respectively) as the sludge age of the system approaches infinity (i.e. no sludge wastage). This has important economic implications. A consequence of low MLSS combined with good organic removal would be a lower energy demand for mixing and oxygen transfer within the bioreactor, in addition to lower sludge disposal costs. It is the combined mixing, scouring and oxygen transfer requirements from the aeration system, which prove to be a significant cost in the submerged MBR system. On the other hand, low MLSS results mainly from long HRTs due to low fluxes, and low fluxes imply high capital costs.

Table 2 Removal of organic and microbial constituents, SidMBR and SubMBR

Determinant	Sub PS MBR	Sub NWPP MBR	Sid EPP
<i>Permeation</i>			
Flux*, l m ⁻² h ⁻¹	7	5	30
Specific flux*, l m ⁻² h ⁻¹ bar ⁻¹	120	83	20
<i>Rejection</i>			
TC, log	~ 8	3-5	~ 5
BOD ₅ , %	97.4	96.9	-
Total COD, %	92.4	93.3	66-82
NH ₃ -N, %	97.6	99.8	~0
Turbidity, %	99.8	99.6	94

*Equilibrium values

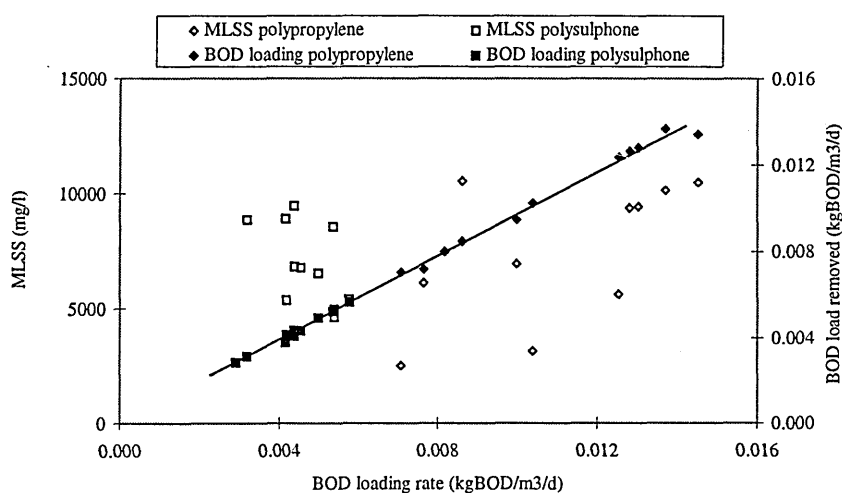


Figure 6 BOD loading rate with respect to load and percentage removal for PS and NWPP membranes

4.0 Conclusions

A comparison between physical treatment under hydrodynamic conditions prevailing in a sidestream MBR and treatment by a submerged MBR, both configurations being based on a low-cost polypropylene membrane material, reveals the material to be extremely susceptible to substantially irreversible fouling.

Physical treatment based on an extruded tubular polypropylene (EPP) material having a pore size of 0.45µm can be improved by cyclic membrane collapsing, though this only achieves partial flux recovery. A kaolin precoat appears to afford the membrane significant protection. Protein fouling, on the other hand, has a profoundly deleterious effect on both flux and, somewhat surprisingly, bacterial rejection.

The submerged configuration based on EPP yields no flux at 0.06 bar (as a hydrostatic head of water). A more open, non-woven hydrophilic polypropylene membrane material (NWPP) yields a very high initial flux which decays to around 8 l m⁻² h⁻¹ after a few hours and gradually fell to around 5 l m⁻² h⁻¹ after a number of weeks. Rejection of bacteria correspondingly increases over this period. Under steady-state conditions the specific flux of the submerged configuration is around 4 times greater than that measured for the sidestream filtration trials.

Membrane fouling of the submerged MBR yields low fluxes and so long hydraulic retention times. This then leads to greater nitrification and low MLSS values - and thus low sludge production. The high membrane area required to maintain throughput, on the other hand, implies high capital costs. Under equilibrium conditions the NWPP MBR yields a similar quality of treated water to the PS MBR, but only once substantially fouled so as to decrease the flux and increase bacterial rejection. The conventional hydrophilic PS membrane, on the other hand, fouls reversibly: full flux recovery is attained after mechanical cleaning. Moreover the equilibrium flux is 60% higher than that of the NWPP membrane.

5.0 Acknowledgements

The authors wish to thank the EPSRC and Severn Trent Water for supporting this work financially, and the Kubota Corporation and Scimat Ltd. for providing the membrane materials.

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CHAPTER FIVE

LOW-COST MEMBRANE MATERIALS FOR USE IN A SUBMERGED MBR: PERMEATE FLUX AND EFFLUENT QUALITY

LOW-COST MEMBRANE MATERIALS FOR USE IN A SUBMERGED MBR: PERMEATE FLUX AND EFFLUENT QUALITY

1.0 Introduction

Municipal wastewater in the UK is often treated by the activated sludge process (ASP) which utilises microorganisms for the natural degradation of pollutants, such as organic carbon and nitrogen compounds. The ASP not only requires large aeration and settlement tanks, it also generates large quantities of excess sludge whose treatment and disposal represents 60% of total treatment costs.[1] In addition, EEC legislation requires that more wastewater receives biological treatment prior to discharge resulting in an increase in the amount of sludge for disposal.[2] A technical improvement to this system is the membrane bioreactor (MBR) which replaces the two stages of the ASP – clarification and settlement – with one, an integrated filtration and clarification step.

Of the two types of MBR that have been developed, the submerged MBR has the major advantage of being more cost efficient as it is usually run utilising the pressure of the mixed liquor to push the permeate through the membrane from the bulk solution.[3] In addition to the low footprint, MBRs also have a low sludge production due to the high biomass concentration and long SRTs. With constant pressure MBRs permeate flux is related not only to transmembrane pressure (head of water) but also to the build-up of fouled materials on the surface of the membrane or within its structure. If membrane fouling occurs then permeate flux will decline, leading to a reduction in the influent flowrate. This has a knock-on effect in relation to the amount of substrate the biomass has for cell maintenance and growth. Bacterial yield depends on specific growth, so a reduction in available substrate leads to a reduction in yield, as the microorganisms utilise the available nutrients for maintenance rather than growth. There is then a corresponding reduction the amount of excess sludge produced.

This paper presents the results from further experimental work, building upon previous work carried out by the authors.[3] The hypothesis behind this set of experiments is that

effluent discharged from traditional MBRs is of extremely high quality, higher than that required by current legislation, and that these membranes are extremely expensive leading to high capital costs of membrane plants for wastewater treatment.[2] If low-cost materials could be substituted for traditional membrane materials, capital costs could fall. These low-cost materials would be likely to produce an effluent of inferior quality than that of traditional membranes, but that this standard would be high enough for current legislation.

2.0 Materials and Methods

Plant configuration and standard analytical methods are the same as in Chapter 3. The materials used in this experimental run were non-woven polypropylene (NWPP) with either a 1.5 or 3 μm pore size (Figures 1,2). These were run alongside a polysulphone membrane with a 0.4 μm pore size (PS); this membrane is currently used in full-scale submerged MBRs for treating municipal wastewater.



Figure 1 Upper surface SEM view of 1.5 μm non-woven polypropylene membrane material

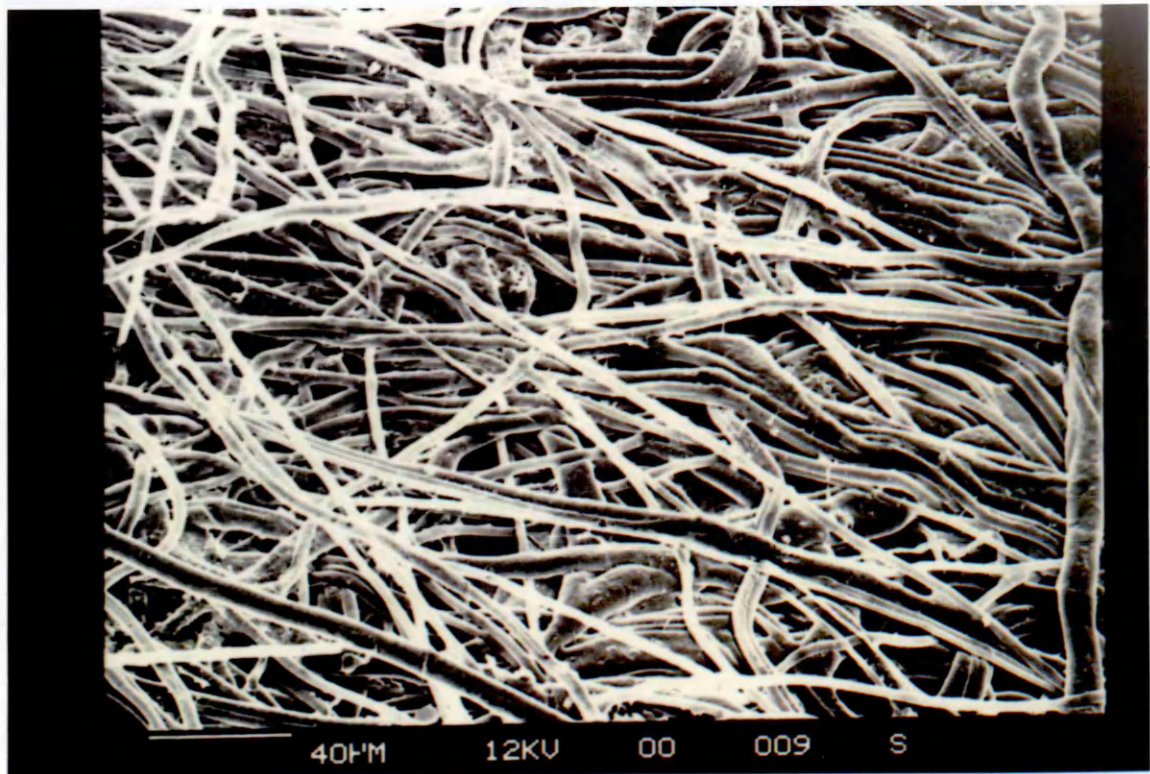


Figure 2 Upper surface SEM view of 3 µm non-woven polypropylene membrane material

3.0 Results and Discussion

3.1 *Clean water flux and initial flux decline*

From Figure 3 it can be seen that the clean water flux from the PS is the lowest of the three membranes at $177 \text{ l m}^{-2} \text{ h}^{-1}$ which does not decline from the beginning to the end of the run. The 3 µm NWPP membrane shows the highest flux followed by the 1.5 µm membrane. The 3 µm NWPP membrane has an initial clean water flux of 225 which falls to $210 \text{ l m}^{-2} \text{ h}^{-1}$ at the end of the run. The 1.5 µm NWPP membrane has an initial flux of 200 which falls to $185 \text{ l m}^{-2} \text{ h}^{-1}$ at the end of the run. All of the membranes have pore sizes that do not pose a barrier to material present in tap water, so the membranes should not become fouled.

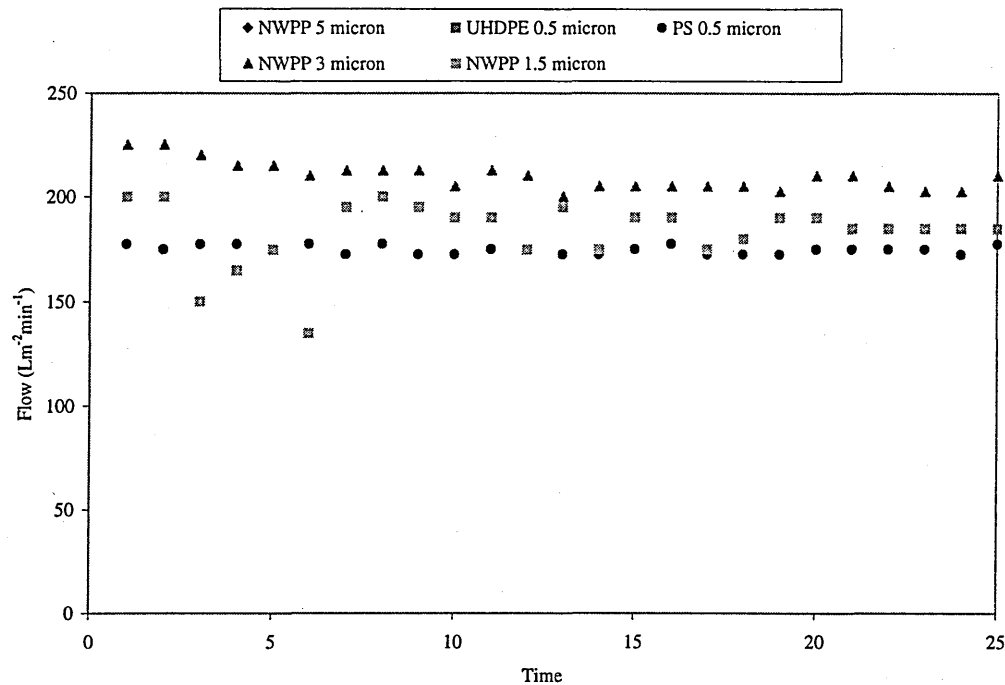


Figure 3 Clean water fluxes from the PS and NWPP membranes

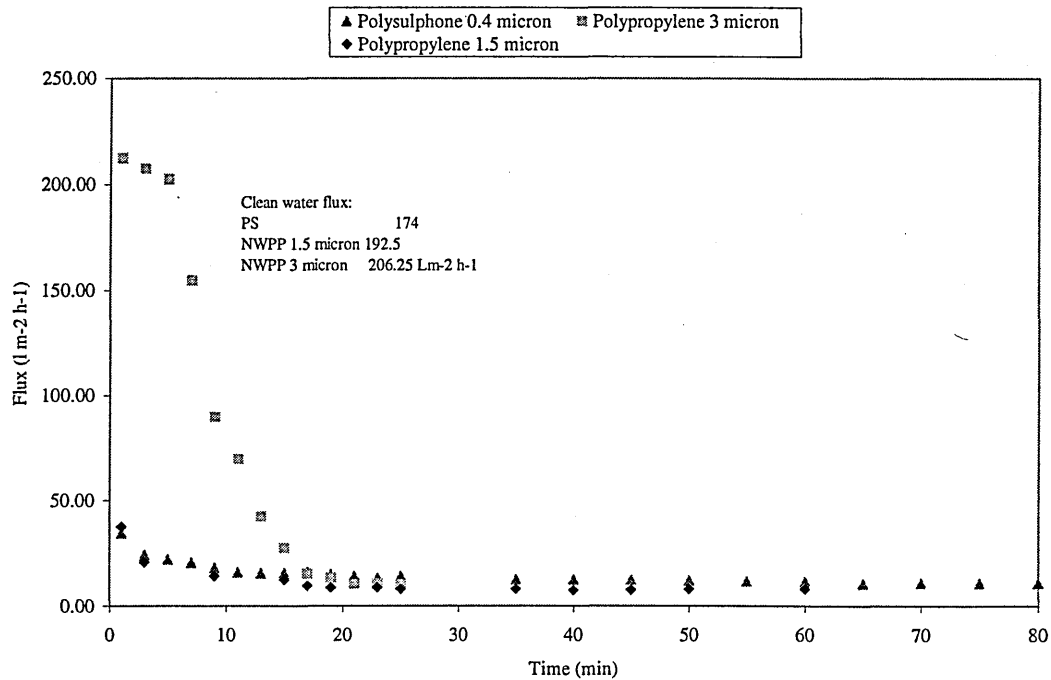


Figure 4 Flux decline from the PS and NWPP membranes in the first 80 minutes from start-up

The bioreactors were seeded with activated sludge on day 0 and were left to acclimatise for 24 hours. On day 1 the permeate drains were opened and the feed flow initialised. As can be seen from Figure 4 there is a rapid initial decrease in flux from the 3 μm NWPP membrane from 212 to $10 \text{ l m}^{-2} \text{ h}^{-1}$. This may be due to the relatively large pore size and the tortuous nature of the pores which may lead to the external and/or internal surfaces being blocked or.[4,5] It is generally recognised that fouling of the membrane is most critical at the start of filtration (or when a change in pressure or flux is most rapid), and the build-up of the fouling layer in the first few hours of start-up is thus critical in determining operability, and is the basis of the critical flux hypothesis.[6,7] The PS membrane that was run alongside this NWPP membrane did not show a very distinct initial decrease in flux. The starting flux when the permeate drains were opened was 52.5 and this only declined to $41 \text{ l m}^{-2} \text{ h}^{-1}$ over the first hour. The 1.5 μm NWPP membrane and the PS membrane which was run alongside showed a very similar flux decline in the first 1.5 hours from ~ 35 down to $10 \text{ l m}^{-2} \text{ h}^{-1}$.

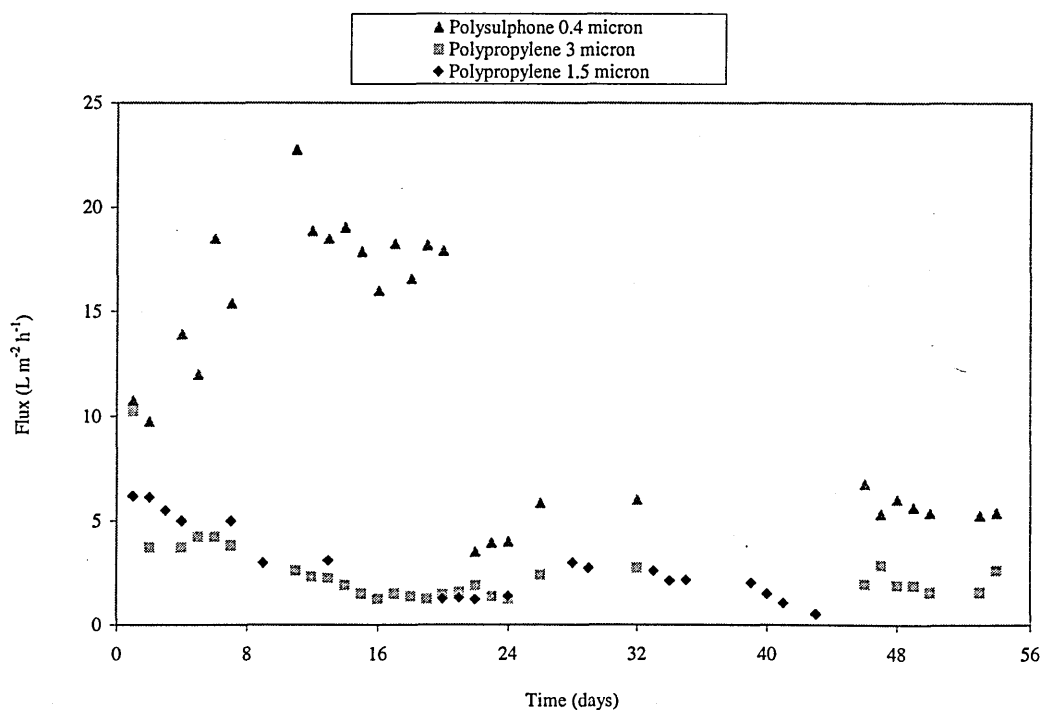


Figure 5 Permeate flux decline over the experimental period for the PS and NWPP membranes

Figure 5 shows the permeate flux decline over the period of the experimental run. It can be seen that the PS membrane, on both runs, had a highly changeable flux. On day 22 the MLSS level in the bioreactor fell to below the top of the membrane. This led to a permanent decrease in the flux from the PS membrane but this flux was then maintained for the rest of the run at $5 \text{ l m}^{-2} \text{ h}^{-1}$. The $1.5 \text{ }\mu\text{m}$ NWPP membrane showed a steady decrease in flux until day 22, when this flux was then in a constant state of change and no steady-state flux was achieved. The $1.5 \text{ }\mu\text{m}$ NWPP membrane showed a similar flux decline to that of the $3 \text{ }\mu\text{m}$ NWPP membrane, and this was again because of the problems with the plant on day 26 of the run.

3.2 Total coliform removal

Figure 6 shows the log reduction of total coliforms for the three membranes. Effluent from the PS membrane showed ~7-log reduction when run with the $3 \text{ }\mu\text{m}$ NWPP membrane. Total coliform removal from the PS membrane run with the $1.5 \text{ }\mu\text{m}$ NWPP membrane was only ~5-log. Both the NWPP membranes showed a 4-log reduction. Chiemchaisri *et al.* showed that membranes with 0.03 and $0.1 \text{ }\mu\text{m}$ pore size attained the same log reduction of coliphage viruses and that improved rejection occurred with time due to the build-up of the dynamic layer.[8] Figure 6 also shows that the 1.5 and $3 \text{ }\mu\text{m}$ NWPP membranes achieve the same log-reduction for total coliforms. From data gathered on day 1 it is also apparent that the build-up of the dynamic layer appears to have some effect on total coliform removal. The PS membrane let 686 bacteria through on day 1, but on day 4, Figure 6 shows that no bacteria were let through the membrane. The $1.5 \text{ }\mu\text{m}$ NWPP membrane let through 52,000 bacteria on day 1 but this had declined to 5810 by day 4. These data show that the effluent from the NWPP 1.5 and $3 \text{ }\mu\text{m}$ pore size membrane would not meet the Bathing Water Directive.[9] This is an important consideration as MBRs are often installed for their disinfection capabilities.

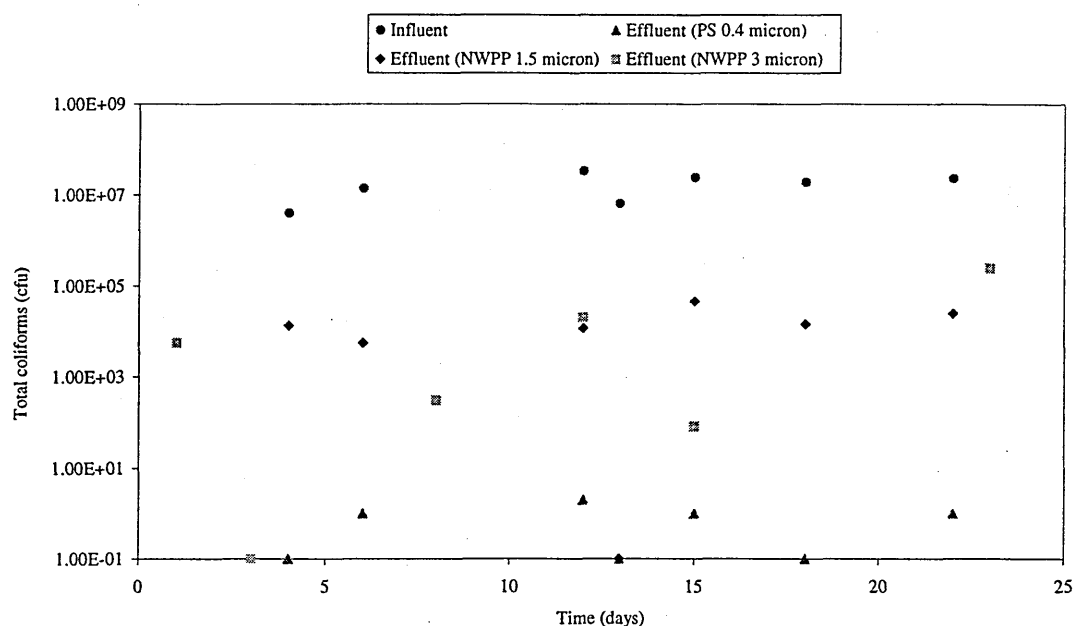


Figure 6 Total coliform removal from the PS and NWPP membranes

3.3 Organics removal

BOD₅ removal is high at >95% for all three membranes. COD removal rates range from 27 to 90%. There was a high variation in influent BOD₅ load, but this did not have an adverse effect on removal rates, even when the MLSS had fallen to below 5000 mg l⁻¹. From Table 1 the BOD:COD ratio can be seen. The BOD:COD ratio of the influent is only 0.6 at its highest, which shows that Cranfield domestic sewage has less recalcitrant organics than normal municipal sewage (BOD:COD ratio of 1.7). The BOD:COD ratio from the effluent from the PS membrane only rises to 0.18, whereas the ratio of the effluent from the 1.5 µm NWPP membrane rises to 0.23 and 3 µm NWPP membrane to 0.34. All these figures show that the MBR is highly efficient in treating the recalcitrant organics.

Table 1 Average BOD:COD ratios

Influent	Effluent PS 0.4 µm	Effluent NWPP 1.5 µm	Effluent NWPP 3 µm
0.37	0.04	0.14	0.08

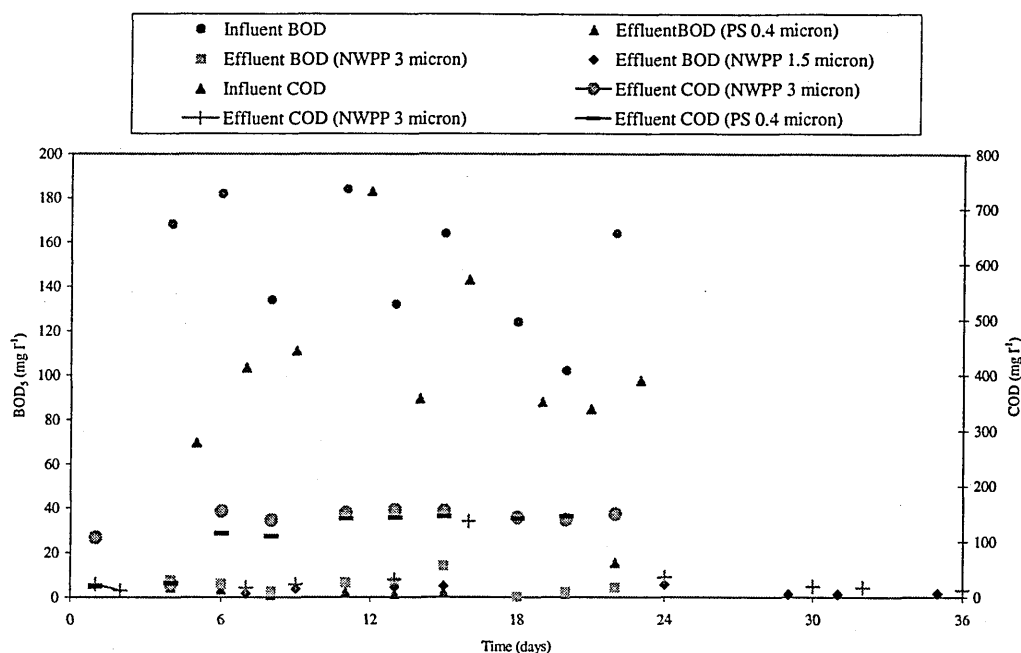


Figure 7 Effluent BOD₅ and total COD concentrations from the PS and NWPP membranes

3.4 Mixed liquor suspended solids

From Figure 8 it can be seen that the MLSS start off high when the bioreactors are seeded with activated sludge on day 0, and then fall over time until an equilibrium value is reached. On day 20 the biomass between the two reactors (3 μm NWPP and PS). The bioreactor containing the PS membrane seemed to reach steady-state at $\sim 5000 \text{ mg l}^{-1}$, whereas the MLSS in the bioreactor containing the 3 μm NWPP membrane did seem to be fall again.

The MLSS in both the bioreactors containing the 1.5 μm NWPP membrane and the PS membrane seemed to fall from their initial high levels. On day 26 the bioreactors half emptied and sewage was used to fill them. On day 28 it can be seen that the MLSS concentration rose, although this started to drop again over the rest of the time period. For all of these mixed liquors no noticeable steady-state value was achieved.

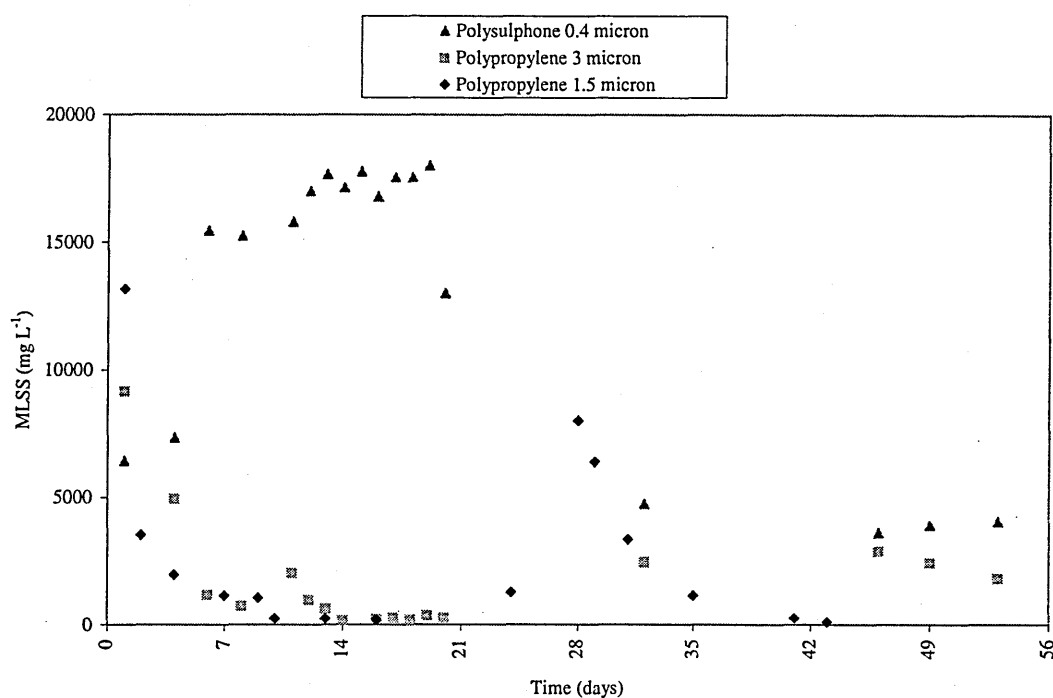


Figure 8 MLSS concentrations in the membrane bioreactors

The high levels of viable bacteria in the original activated sludge that the bioreactors were seeded with could not be expected to be maintained at such low permeate fluxes – leading to low feed flows. An equilibrium concentration of viable bacteria should have been achieved from the available substrate concentration. In substrate-limited wastewater processes such as MBRs, bacteria will preferentially allocate the available carbon source toward satisfying their maintenance energy requirements. It has been reported by Low and Chase that a reduction in the biomass production will result in less nitrogen being removed by assimilation, although this has not been proved in this work, where ammonia removal was always higher than 60% and more usually over 95% (Table 2).[10]

Table 2 Ammonia removal from the effluents from the PS and NWPP membranes

Day	NH ₃ removal (%)		
	PS	3 μ m NWPP	1.5 μ m NWPP
1	100	98	99
2	99	99	63
7	93	66	100
9	100	83	88
13	99	26	57
24	100	12	37
29	99	100	91
31	100	100	88
35	89	96	100

4.0 Conclusions

- The NWPP 3 μ m pore size membrane showed a drastic reduction in permeate flux in the first 10 minutes of filtration.
- All the membranes tested showed a drastic flux decline over time.
- Both the NWPP and PS membranes produced an effluent extremely low in organic matter well within the current legislation requirements even when the MLSS concentrations fall in the bioreactors.
- Ammonia removal is still high with low MLSS.
- The PS membrane can achieve a 9-log reduction in total coliforms, whereas the NWPP membranes can only achieve a 4 or 5-log reduction.
- The build-up of the dynamic layer increases bacterial rejection.
- Low BOD:COD ratio shows MBRs are good at treating recalcitrant organics.
- MBRs can accommodate natural variations in load.
- The robustness of MBRs is indicated by their ability to absorb natural variations in load without an adverse effect on treatment capability.
- The ability of MBRs to produce a high quality effluent even when MLSS were low may lead to zero sludge disposal costs, due to the available substrate being utilised for bacterial cell maintenance and not growth.

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CHAPTER SIX

CONCLUSIONS AND RECOMMENDATIONS

1.0 Introduction

The present work aims to assess the feasibility of using low-cost membrane materials in submerged membrane bioreactors (MBRs) to treat domestic wastewater. Capital costs of MBRs are high, due mostly to the expense of the membrane, therefore, low-cost membranes would reduce the capital costs, although membrane area may have to increase. The effects of membrane fouling and effluent quality were investigated with respect to specific flux, bacteria and organics removal.

1.1 Conclusions

1.1.1 MBRs

Reduced sludge production

Long sludge retention times (SRTs) lead to endogenous respiration within the biomass and, therefore, reduced sludge production. Long SRTs lead to the growth of specialised bacteria, e.g. nitrifiers, and enhanced breakdown of recalcitrant organics. Organic loading rates are restricted by flux but are generally higher than conventional activated sludge plants with greater carbon and nitrogen removal rates. Substrate inhibited processes such as MBRs lead to reductions in viable bacteria concentrations due to the requirement to use all available nutrients for cell maintenance rather than growth. Long SRTs and reduced biomass production leads to a reduced sludge production and therefore, sludge disposal costs.

Operating costs

Aeration costs are high in submerged MBRs due to the need to scour the membranes, set up the crossflow effect and to provide oxygen to the biomass. This often leads to the requirement for coarse and fine bubble aerators, compared to only fine bubble aerators in sidestream configurations. Aeration in submerged MBRs can be >90% of the total power requirements compared with 20% in sidestream systems. Aeration costs in MBRs may be reduced by running the systems at lower MLSS concentrations leading to a lower oxygenating and mixing demands.

Submerged systems have very little pumping costs unless permeate flux is enhanced with a suction pump, which can contribute up to 28% of the total costs. However, low

costs are associated with low fluxes compared with sidestream systems, which in turn increases capital costs and footprint area. Sidestream systems have liquid pumping costs of 60–90% of the total costs, and this along with the high cleaning requirements leads to much higher operating costs compared with submerged systems, although the capital costs and footprints are smaller, as the membrane area required is smaller.

Sidestream systems have regular and intensive cleaning regimes, to reduce membrane fouling. Membranes in submerged MBRs have long running periods (up to a year) with no cleaning requirement, due to the much lower permeate fluxes. Cleaning is then generally much easier without the need for downtime.

1.1.2 Membranes

Bacterial rejection

Significant bacteria and virus rejection is accomplished with MF and UF membrane filtration. Rejection with MF membranes is enhanced by the build-up of a dynamic layer. In general the PS membrane rejected 99.9% of total coliforms whereas the NWPP membranes rejected approximately 90%. This impacts on current legislation requirements for effluent discharge to bathing waters—the PS membrane would meet the requirements but the NWPP membranes would not. This is significant as MBRs for municipal wastewater treatment are generally used for their disinfection capabilities.

Flux decline

Rejected matter accumulates at the membrane surface as a concentration polarisation layer that leads to a resistance to filtration and membrane fouling. Flux decline has been shown to be important in the early stages of filtration and the importance generally declines with time. The NWPP membranes showed a drastic decline in flux in the first 10 minutes of filtration. The PS membrane also declined until it reached a steady-state flux where no further decrease was apparent. This may be attributable to the constant flux hypothesis.

Organics removal

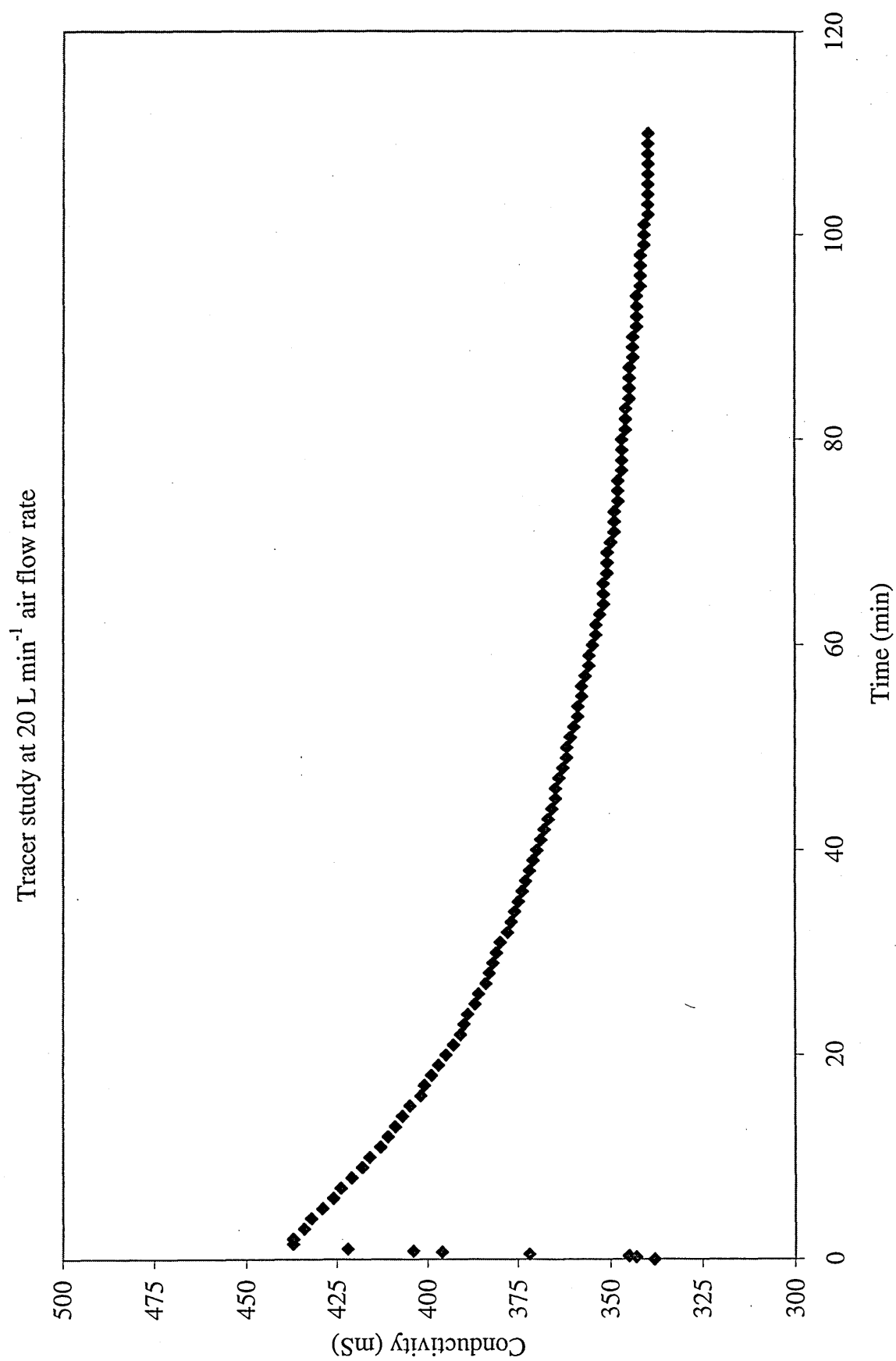
Long SRTs lead to breakdown and assimilation of organic matter including recalcitrant organics. This leads to effluents having extremely low BOD and COD concentrations. Even when the MLSS concentrations have fallen to $<5000 \text{ mg l}^{-1}$, BOD_5 removal is still $>60\%$. The robustness of the MBR is indicated by its ability to absorb natural variations in load.

Although the NWPP membranes are relatively inexpensive to manufacture compared with the PS membrane, their susceptibility to fouling and poor rejection means that they would be inappropriate for use in a submerged MBR treating domestic wastewater.

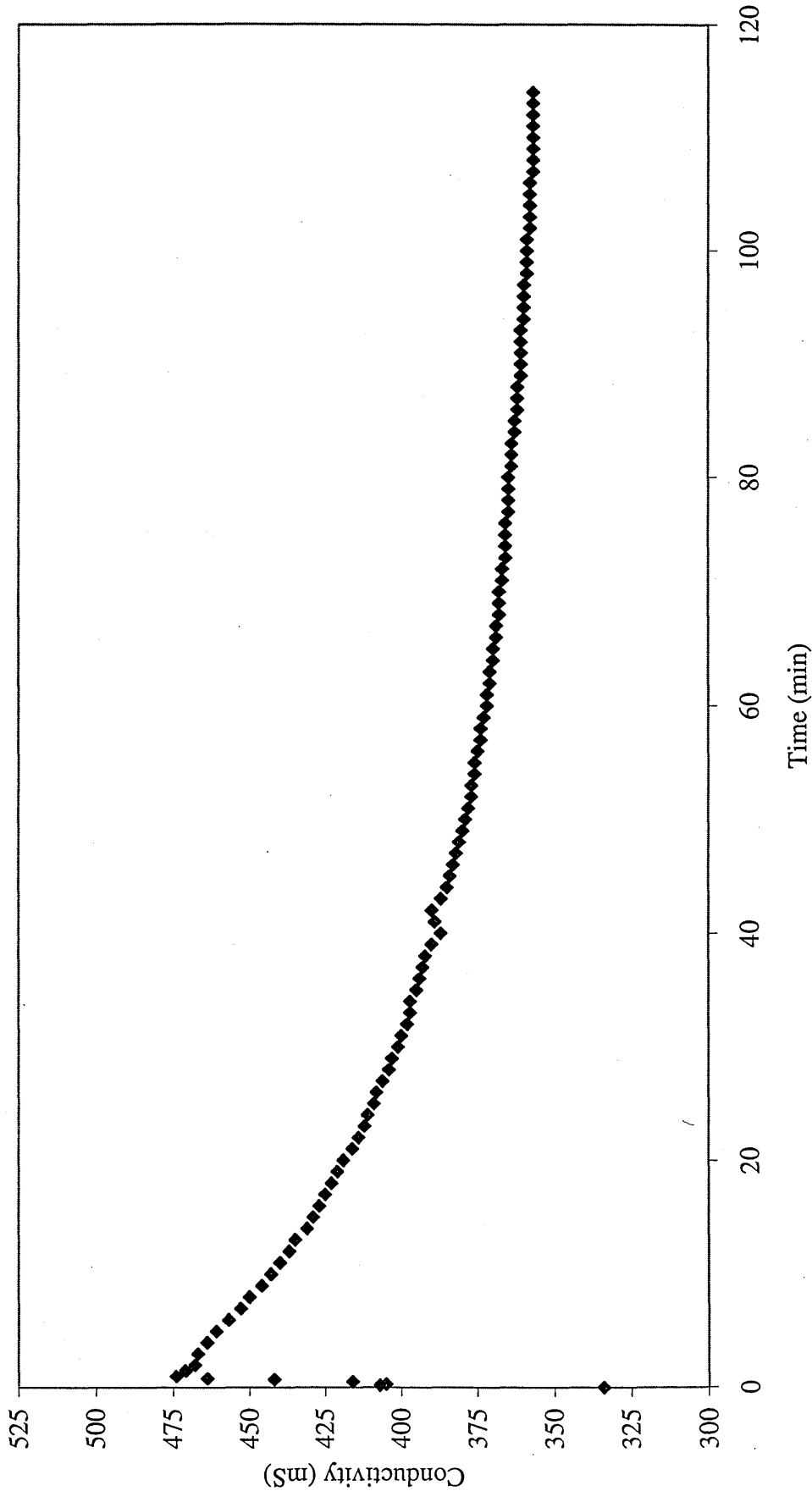
1.2 Recommendations for further work

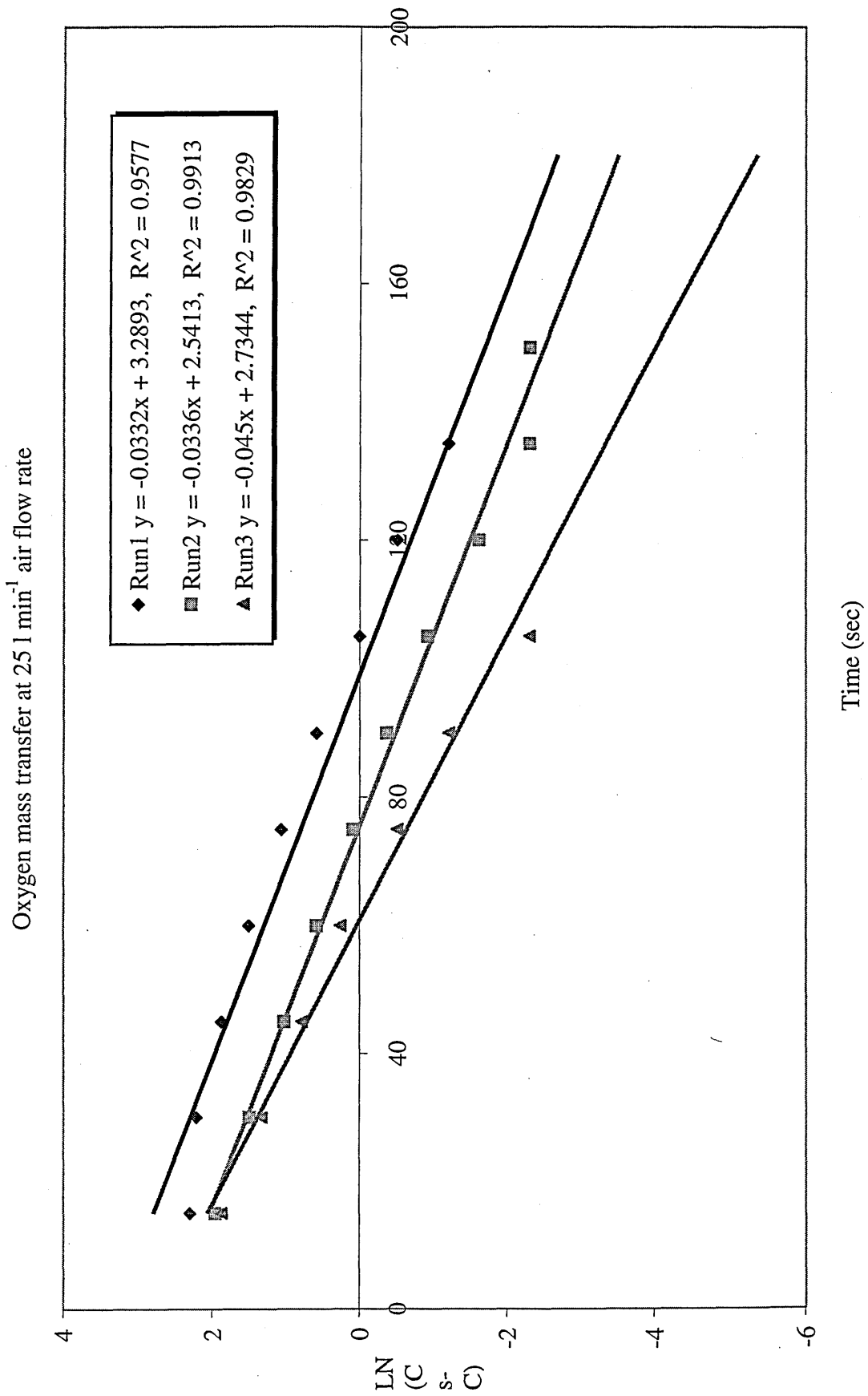
A more in-depth analysis of membrane processes and configuration is required in order to identify the relative merits of submerged and sidestream configurations. The influence of the key process operating conditions and the influence of feedwater quality should be investigated.

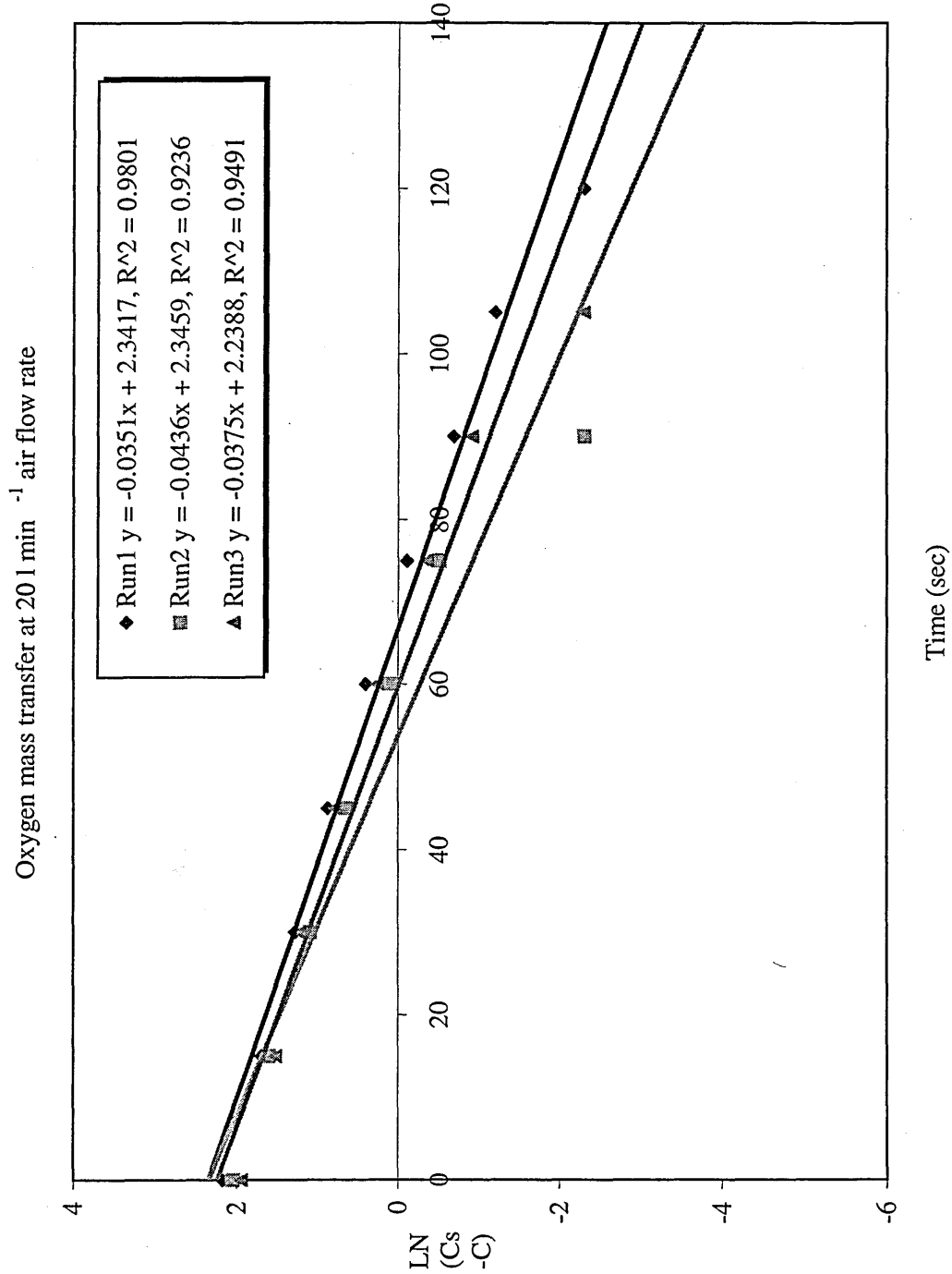
APPENDICES

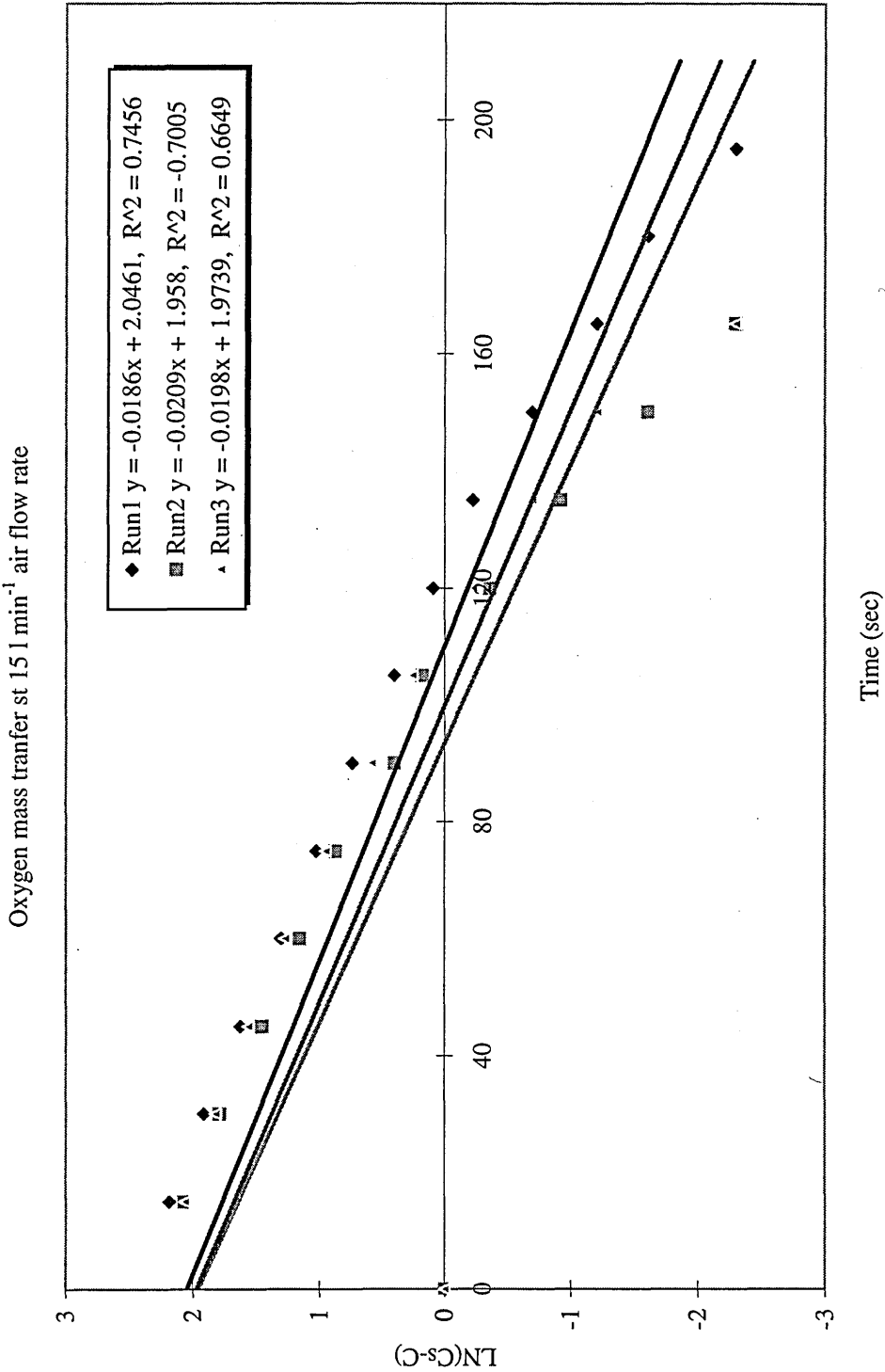


Tracer study at 5 l min⁻¹ air flow rate









Oxygen mass transfer at 10 l min⁻¹ air flow rate

